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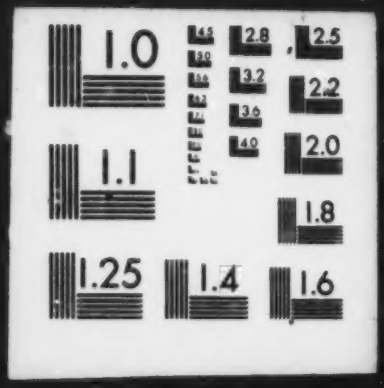

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MOLECULAR SIEVE CONTROL PROCESS IN SULFURIC ACID PLANTS

D. W. Hissong

Battelle Columbus Laboratories

Prepared for:

Industrial Environmental Research Laboratory

October 1975

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ABSTRACT

An engineering analysis of the applicability of molecular sieve technology to the control of sulfur dioxide emissions from sulfuric acid plants has been conducted. Field test data from a commercial plant using this technology show that after the equivalent of 10 months of operation the plant is still controlling the sulfur dioxide emissions to well within the existing Federal and state regulations for sulfuric acid plants. This plant is also meeting the performance guarantee of the process developer and vendor (Union Carbide Corporation). The average emission after 10 months was 0.9 lb SO_2/ton acid, compared with a most stringent existing regulation of 4.0 lb SO_2/ton acid and a performance guarantee equivalent to about 1.2 lb SO_2/ton acid. The concept of a 2-year sieve life with acceptable SO_2 control has not been demonstrated, although there is no reason to believe it cannot be achieved. Thus, at this point this application of molecular sieve technology appears technically feasible.

The economic feasibility of molecular sieve technology for this application was assessed by comparing the total capitalized cost (including investment and operating cost) of this technology with those of the Wellman-Lord and dual absorption processes. The capitalized costs for the three processes are fairly close, and individual acid plant characteristics will affect the economic choice. The molecular sieve technology is more competitive for the smaller acid plants and for acid plants which already have facilities for drying the air needed to regenerate the sieve. Although for many acid plants the dual absorption process will be the least expensive, this process is limited in its effectiveness such that for most plants it cannot meet the regulation of 4 lb SO_2/ton acid. Thus, considering the overall cost and effectiveness, the molecular sieve technology appears economically feasible for some acid plants.

UNITED STATES ENVIRONMENTAL PROTECTION AGENCY
NATIONAL ENVIRONMENTAL RESEARCH CENTER
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FROM: Gilbert M. Gigliotti G.M.G./J.B.
Director, Technical Information Staff

TO: Input Branch
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Springfield, Virginia 22161

Enclosed are two copies of "Vapor-Phase Organic Pollutants - Volatile Hydrocarbons and Oxidation Products" by Panel on Vapor-Phase Organic Pollutants (EPA-600/1-75-005) and one copy of "Molecular Sieve Control Process in Sulfuric Acid Plants" by D.W. Hissong (EPA-600/2-75-066) for inclusion in the National Technical Information Service. All required information and forms are included.

If you have any questions or need any other information, please call Mrs. Thomasine Bayless (513/684-7282). Thank you for your cooperation.

3 Enclosures

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ENGINEERING ANALYSIS OF A MOLECULAR SIEVE CONTROL PROCESS IN SULFURIC ACID PLANTS

by

D. W. Hissong

INTRODUCTION

The manufacture of sulfuric acid results in emissions of sulfur oxides and sulfuric acid mist. The emissions of acid mist can be reduced to meet the anticipated regulations through the use of efficient mist eliminators. A number of technologies are being considered for controlling the sulfur oxide emissions. Among the alternatives is a process involving adsorption of the sulfur oxides onto a molecular sieve followed by cyclic regeneration of the adsorbent and recovery of the sulfur oxides. This process is called the PuraSiv S process and has been developed by Union Carbide Corporation. Union Carbide claims that this process can reduce the average concentration of sulfur oxides in the effluent gas stream to less than 100 ppm by volume, which corresponds to an emission rate well below the present or anticipated regulations.

In an effort to determine the technical and economic feasibility of molecular sieve control systems for this application, the U.S. Environmental Protection Agency sponsored this study. The objectives were to determine

- The technical feasibility of this control system by analyzing data independently obtained from an operating unit as well as information obtained from the system developer
- The economic feasibility of the system by comparing it with other commercially available control processes.

SULFURIC ACID MANUFACTURING PROCESSES

Nature of the Industry

Sulfuric acid is a very basic industrial chemical, being involved in countless manufacturing operations. It is of such paramount significance

that the consumption of this chemical has been used as a business indicator and a measure of technical development of nations. The production of sulfuric acid in the United States in 1973 was 28.8 million metric tons.[1] About half of this acid was used in the manufacture of phosphate fertilizers.

Essentially all the sulfuric acid in the United States is now produced by the Contact Process, less than 0.4 percent being produced by the older Chamber Process. There are currently 152 plants in this country using the Contact Process and only five plants using the Chamber Process.

The Contact Process

A flow sheet for the Contact Process is shown in Figure 1.[2] The source of sulfur for the process can be elemental sulfur, spent sulfuric acid (sludge), hydrogen sulfide, or other sulfur compounds. This source is burned with dry air in a furnace to produce sulfur dioxide (SO_2). The latter is further oxidized to sulfur trioxide (SO_3) over a catalyst which usually consists of supported vanadium pentoxide. The resulting gases are cooled and then introduced into an absorption tower in which the SO_3 is absorbed in water to form sulfuric acid. The reactions involved are as follows (writing the sulfur source as S).

Burner $S + O_2 \rightarrow SO_2$

Oxidation reactor $SO_2 + 1/2\ O_2 \rightarrow SO_3$

Absorption tower $SO_3 + H_2O \rightarrow H_2SO_4$

The product is usually marketed as 93-98 percent H_2SO_4 solutions or oleums, which are solutions of SO_3 in H_2SO_4.

Contact sulfuric acid plants are usually classified as either "sulfur burning plants" or "wet gas plants". However, the significant distinction for this study is between the wet gas plants and what one might call "dry gas plants". About three-fourths of the sulfuric plants in this country are of the dry type, which means that the gas leaving the burner is kept as dry as possible by drying the input streams to the burner. This drying is done with some of the concentrated acid (98 weight percent) and is necessary if the plant is to produce a concentrated acid product.

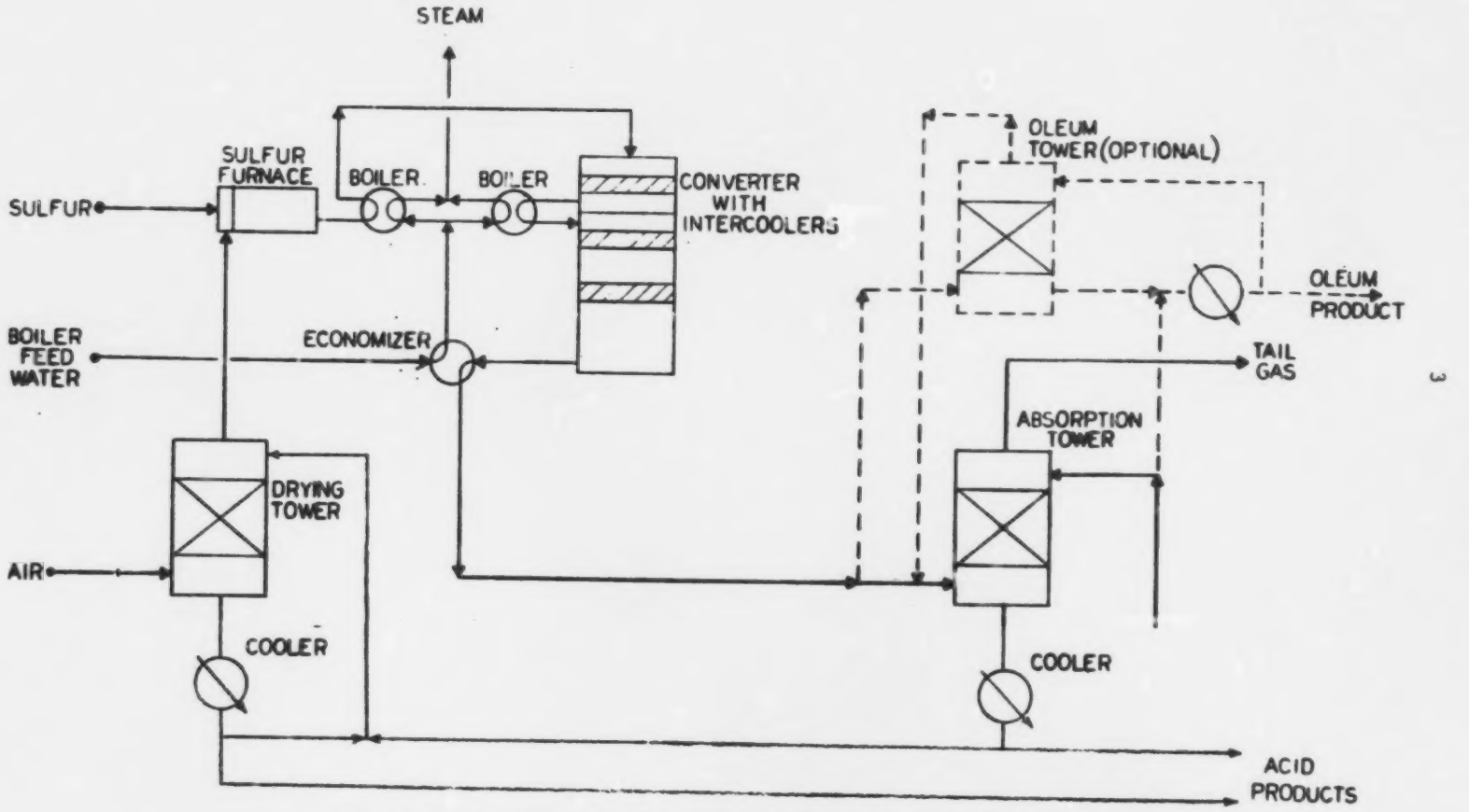

FIGURE 1. FLOW SHEET OF SULFUR-BURNING CONTACT SULFURIC ACID PLANT

All plants which feed exclusively elemental sulfur or sulfur ores are operated dry. However, plants feeding a hydrogen sulfide stream can be operated either wet or dry. There are two consequences of wet operation which are important here

- The volume of the tail gases to be treated is greater for wet gas plants than for dry gas plants
- Wet gas plants do not have available a source of dried air for regenerating the molecular sieve.

Plant Size and Type Distribution

Of the 152 contact sulfuric acid plants in the United States, 112 are dry gas plants and 40 are wet gas plants. The total capacity of the dry gas plants is 74,310 tons* (67,400 metric tons) per day and that of the wet gas plants is 26,205 tons (23,770 metric tons) per day. Table 1 shows the size distribution for both types of plants.[3] The plant capacities range from 15 to 2000 tons per day, with the average capacity being about 660 tons per day.

The 1973 production rate of 31.8 million tons (28.8 million metric tons) per year corresponds to about 87 percent of the capacity listed in Table 1.

Industry Growth Projection

The annual production of sulfuric acid in the United States is projected to increase to about 31 million metric tons in 1976 and 37 million metric tons in 1980.[3] These projections are based upon the assumptions that

- The production of acid from smelter gases remains constant at the present level of 3.4 million metric tons per year
- The fertilizer uses of sulfuric acid increase at a rate of 6 percent per year and the nonfertilizer uses at 4 percent per year.

---

* "Tons" refers to the short tons (2000 lb) normally used in this industry. A "metric ton" is 2,205 lb.

TABLE 1. DISTRIBUTION OF CONTACT SULFURIC ACID PLANTS BY SIZE AND TYPE[a]

Tons per Day		Number of Plants of Each Type		Percent of Capacity of Category[b]	
Size Range	Average Size for Range	Dry Gas	Wet Gas	Dry Gas	Wet Gas
15- 165	106	15	5	2.1	2.5
170- 225	200	13	2	3.5	2.4
245- 285	260	6	5	2.1	4.9
300- 430	360	19	11	8.9	15.8
450- 570	500	14	4	10.0	7.6
600- 615	600	4	2	4.9	4.6
645- 750	700	7	4	6.4	10.7
- 800	800	--	2	--	6.1
- 830	830	1	--	1.1	--
900- 950	900	9	--	10.9	7.1
1000-1040	1015	2	1	2.7	3.8
-1200	1200	1	--	1.6	--
1300-1450	1400	5	1	9.4	5.5
1460-1550	1500	4	--	8.1	--
-1600	1600	6	1	12.9	6.1
1750-1900	1850	4	--	10.0	--
-2000	2000	2	2	5.4	22.9
		112	40	100.0	100.0

(a) Plants existing in 1974 that went on-stream prior to 1973.

(b) Total capacity in dry process plants is 74,310 tons per day; for wet process plants it is 26,205 tons per day.

Emission Sources

The major source of emissions from contact sulfuric acid plants is the exit gas, or tail gas, from the absorber.* This gas contains unreacted SO_2, unabsorbed SO_3, and sulfuric acid mist. In the atmosphere the SO_3 will be converted into sulfuric acid mist by hydrolyzing with moisture in the air. The most abundant pollutant in the tail gas is SO_2.

The amount of SO_2 in the tail gas depends upon the conversion efficiency of the oxidation reactors, which in turn depends on a number of factors including

- Concentration of sulfur dioxide in the gases entering the converter and the ratio of oxygen to sulfur dioxide particularly in the last converter stage
- Number of converter stages
- Volume and distribution of catalyst in various converter stages
- Catalyst efficiency (depends on catalyst type and age)
- Uniformity of gas composition
- Impurities in the entering gas
- Temperature control at various points in the converter (this depends in part on having properly sized inter-stage gas cooling equipment).

Figure 2 shows the relationship of the conversion efficiency to the SO_2 emission in pounds per ton of acid** produced.[4] Most contact sulfur acid plants are purchased with guarantees of 96 to 98 percent conversion efficiency. This corresponds to emissions of about 25 to 55 pounds of SO_2 per ton of acid. For comparison, the New Source Performance Standard is 4 pounds of SO_2 per ton of acid.[5]

Figure 3 shows the relationship of the conversion efficiency to the SO_2 concentration in the tail gas based on 26 commercial acid plants.[4] The average for these plants was about 97 percent conversion efficiency and an SO_2 concentration of about 2700 ppm in the tail gas. Table 2 summarizes the data on which this Figure was based.

Table 2 also summarizes the concentrations and emission rates of sulfuric acid mist for the plants. These values vary widely depending on details of the design and operating conditions. Many plants which do

---

* The other sources of atmospheric emissions include gases vented from storage tanks, tank cars, and tank trucks during filling operations and from acid concentrators. Although few data are available on emissions from these sources, they are evidently at least an order of magnitude less than the tail gas emissions.

** kg SO_2/metric ton acid = (0.5)(lb SO_2/ton acid).

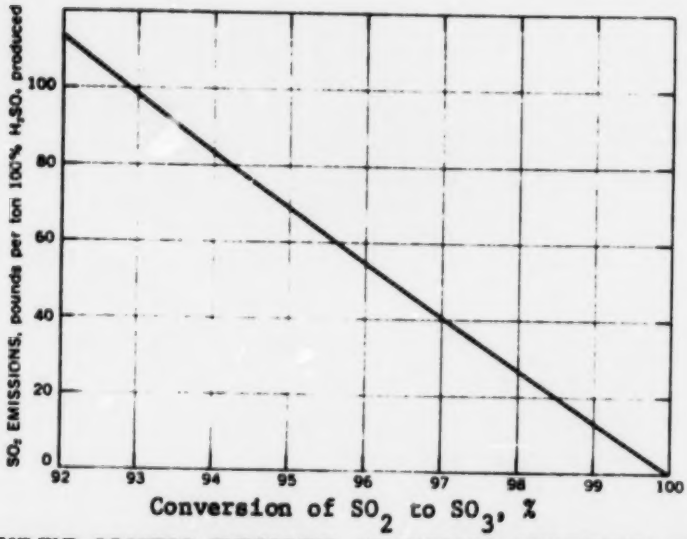

FIGURE 2. SULFUR DIOXIDE EMISSIONS AT VARIOUS CONVERSION EFFICIENCIES (Per Ton of Equivalent 100% H_2SO_4 Produced)

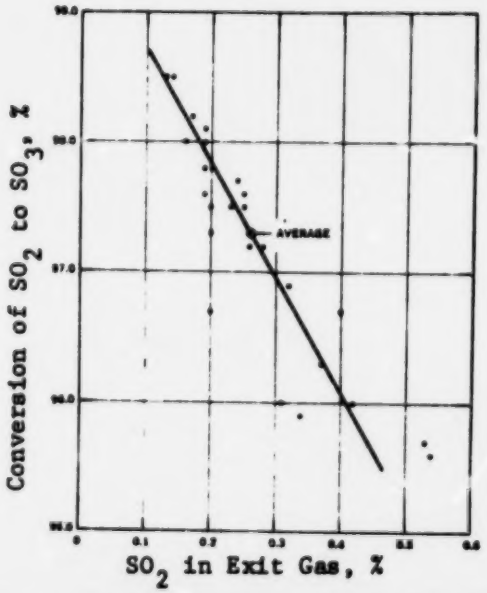

FIGURE 3. RELATIONSHIP OF CONVERSION EFFICIENCY TO SO_2 CONCENTRATION IN EXIT GASES

TABLE 2. CONTACT SULFURIC ACID PLANT TAIL GAS EMISSIONS

Type of Plant		Controls	Tail Gas Rate (10^3 scf/T)	SO_2 Conc. (volume percent)	SO_2 Emission (lb/T)	Mist Conc. (mg/scf)	Mist Emission (lb/T)
Sulfur burning							
Air dilution	-1	None	88.2	0.20	31.3	2.4	0.46
	-2	↓	106.3	0.31	58.5	19.6	4.6
	-3		91.2	0.23	36.7	3.0	
	-4		85.6	0.25	41.5	9.2	0.57
	-5		83.1	0.28	41.5	9.2	1.7
	-6		93.6	0.20	34.0	3.0	0.80
No air dilution	-1	None	73.8	0.53	70.2	3.7	0.58
	-2		73.9	0.19	24.3	2.3	0.35
	-3		82.3	0.24	35.2	9.5	1.7
	-4		86.4	0.40	61.6	6.1	1.2
	-5		79.0	0.25	34.8	1.9	0.32
	-6	↓	98.9	0.42	56.6	37.3	6.3
Air dilution		Mist Elim.	86.9	0.14	21.6	6.5	
No air dilution		Mist Elim.	71.0	0.19	24.7	13.5	
Average			85.7	0.27	40.6		
Combination							
Spent Acid H_2S S							
X X	-1	None	129.6	0.37	85.2	8.4	2.4
X X	-2	↓	93.9	0.20	33.1	11.0	2.3
X	-3		99.2	0.34	60.0	10.2	2.2
X X X	-1	Mist Elim.	78.9	0.32	36.5	0.3	0.03
X X X	-2		75.8	0.17	22.6	0.3	0.05
X X X	-3		81.6	0.32	46.7	0.5-2.0	0.07-0.3
X X X	-4		88.4	0.18	27.6	26.3	0.22
X X	-5		102.2	0.20	36.8		
X X	-6	↓	158.4				

Note: scf = cubic feet at 32 F and 1 atm; T = short ton of 100 percent H_2SO_4.

not have an external mist eliminator do have some mist elimination facilities built into the top of the absorber. Also, impingement in the stack and lines leading to it may reduce the mist content of the gas before it is exhausted to the atmosphere. The mist emission rates for the sulfur burning plants listed in Table 2 vary from 0.3 to 6.3 pounds per ton of acid produced. For comparison, the New Source Performance Standard is 0.15 pounds per ton of acid.(5)

Because the sampling and analytical techniques for SO_3 are more complex than for SO_2 and acid mist, the test data on SO_3 are more limited. For the plants in Table 2 for which SO_3 concentrations were reported the values ranged from 0.5 to 48 milligrams per cubic foot. This SO_3 ultimately adds to the total acid mist emission of the plant. The range of concentrations just cited corresponds to contributions of 7 to 85 percent of the total acid mist emissions from the plants involved. Thus, the unabsorbed SO_3 can comprise an appreciable part of the total acid mist emission.

In the absence of further controls, the emission of SO_2 from sulfuric acid plants, excluding smelter gas plants, is estimated to be about 630,000 metric tons in 1976 and 764,000 metric tons in 1980. The corresponding emissions of acid mist are 19,000 metric tons in 1976 and 24,000 metric tons in 1980.(3)

The volume of the tail gas stream is important in designing emission control equipment. One can see from Table 2 that the average tail gas rate for sulfur burning plants was 85,700 scf per ton of acid produced. This value was used for the dry gas plants. For wet gas plants the tail gas rate per ton of acid is about 1.5 times that for dry gas plants.(3)

For wet gas plants the SO_2 concentration in the tail gas is about 3500 ppm compared to the previously quoted 2700 ppm for dry gas plants.(3) This higher concentration plus the higher flow rate means that the SO_2 emission per unit of acid produced is about twice that for dry gas plants.

Emission Regulations

As mentioned previously, the New Source Performance Standards for sulfuric acid plants limit the emissions to 4 pounds of SO_2 and 0.15 pounds of acid mist per ton of acid produced (2 kg of SO_2 and 0.075 kg of acid mist per metric ton of acid). These limits are based on 2-hour

average conditions and apply to plants where the construction or modification started after August 17, 1971. They do not apply to smelter gas plants and other situations in which the acid is made primarily as a means of controlling SO_x pollution.

For sulfuric acid plants built before August 17, 1971, the state regulations apply. Table 3 lists the regulations proposed in the State Implementation Plans. One can see that most states limit the SO_2 emissions to 6.5 pounds per ton of acid and the acid mist emissions to 0.5 pounds per ton of acid.

TECHNICAL EVALUATION OF MOLECULAR SIEVE CONTROL TECHNOLOGY

General Description of Molecular Sieves

The term "molecular sieve" was originated by J. W. McBain to describe fine porous solid materials which have a structure of channels and voids which permits them to act as sieves on a molecular scale. A molecular sieve has pores of a very uniform size which are uniquely determined by the unit structure of the crystal. These pores will completely exclude molecules which are larger than their diameter.

An important type of molecular sieve is the zeolites, which are crystalline, hydrated aluminosilicates of the alkali, and alkaline earth elements. The structure of zeolites consists of a three-dimensional network of SiO_4 and AlO_4 tetrahedra linked together by sharing all the oxygen atoms. There are 34 known species of natural zeolites and about 100 types of synthetic zeolites, although only a few have practical significance at the present time.[6]

The selectivity of a molecular sieve for a particular adsorbate depends not only on the dimensions of the interstitial spaces in the sieve crystal, but also on the polarity of the adsorbate, and, in the case of hydrocarbons, the extent of carbon-bond saturation of the adsorbate.

Use of Molecular Sieves for SO_2 Recovery

In 1963 the Bureau of Mines reported on a study on the use of molecular sieves for recovering SO_2 from industrial gases.[7] The impetus for this study included both air pollution abatement and supplementing production of sulfur from "secondary" mineral sources such as gypsum. A number of

TABLE 3. STATE EMISSION REGULATIONS FOR SULFURIC ACID PLANTS

	Maximum Emission, lb/ton 100 percent H_2SO_4	
	SO_2	Acid Mist
Alabama	6.5	0.5
Alaska	6.5	0.5
Arizona	6.5	0.5
Arkansas	6.5	0.5
California	6.5	0.5
Colorado	6.5	0.5
Connecticut	6.5	0.5
Delaware	14.2	0.5
District of Columbia	6.5	0.5
Florida	4.0	0.15
Georgia	10.0	0.15
Hawaii	6.5	0.5
Idaho	6.5	0.5
Illinois	40.0	0.15
Indiana	6.5	0.5
Iowa	6.5	0.5
Kansas	30.0	0.5
Kentucky	6.5	0.5
Louisiana	6.5	0.5
Maine	6.5	0.5
Maryland	6.5	0.5
Massachusetts	27.0	0.5
Michigan	6.5	0.7
Minnesota	6.5	1.7
Mississippi	6.5	0.5
Missouri	6.5	0.5
Montana	6.5	0.5
Nebraska	6.5	0.5
Nevada	6.5	0.5
New Hampshire	6.5	0.18
New Jersey	6.5	1.88
New Mexico	40.0	0.5
New York	6.5	0.5
North Carolina	27.0	0.5
North Dakota	6.5	0.5
Ohio	6.5	0.5
Oklahoma	4.0	0.15
Oregon	6.5	0.5
Pennsylvania	6.5	0.5
Rhode Island	6.5	0.5
South Carolina	4.0	0.5
South Dakota	6.5	0.5
Tennessee	6.5	0.5
Texas	30.0	0.5
Utah	8.0	1.6
Vermont	6.5	0.5
Virginia	27.0	0.9
Washington	6.5	0.5
West Virginia	30.0	0.5
Wisconsin	4.0	0.5
Wyoming	4.0	0.15

synthetic zeolites were tested as selective adsorbents for recovering SO_2 from various synthetic gases and from gases produced by decomposing gypsum. Three zeolites were found which had capacities considered adequate for this application. These capacities were up to 29 grams of SO_2 per 100 grams of adsorbent. Regeneration studies showed that over 90 percent of the SO_2 desorbed from the loaded beds was of sufficient purity for direct liquefaction. However, all the zeolites tested lost adsorptive capacity with repeated use.

The Union Carbide Corporation has been working on the use of molecular sieves for SO_2 removal for over 5 years. This work has included laboratory determinations of SO_2 selectivities and equilibrium loadings, mass transfer rate measurements, and cyclic life tests. This work led to a pilot scale test program and thereby to the development of the PuraSiv S Process to be considered here.

Description of PuraSiv S Process*

The PuraSiv S Process, developed by Union Carbide Corporation, is a fixed bed adsorption system for concentrating and removing SO_2 from sulfuric acid plant tail gas. The adsorbent used is a synthetic molecular sieve which Union Carbide has developed for this purpose. Upon regeneration of the molecular sieve, the SO_2 is recovered in concentrated form and is recycled to the acid plant. Theoretically, the recovered SO_2 should contribute a 2-3 percent increase in acid production rate, although this has not yet been verified in the commercial test unit. The process also removes and recovers SO_3 and acid mist from the tail gas.

Regeneration of the loaded adsorbent bed is accomplished by a thermal swing, using hot air to flush out the SO_2 as a concentrated stream. There are two versions of the process which differ in the source of the air for regeneration. If a source of acid-dried air is available, such air will be used for the regeneration and the process flow sheet will be as in Figure 4. If no such source is available, the regeneration gas will be supplied by drying ambient air with an additional molecular sieve. The flow

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* Based on information obtained from Union Carbide Corporation and Reference 8.

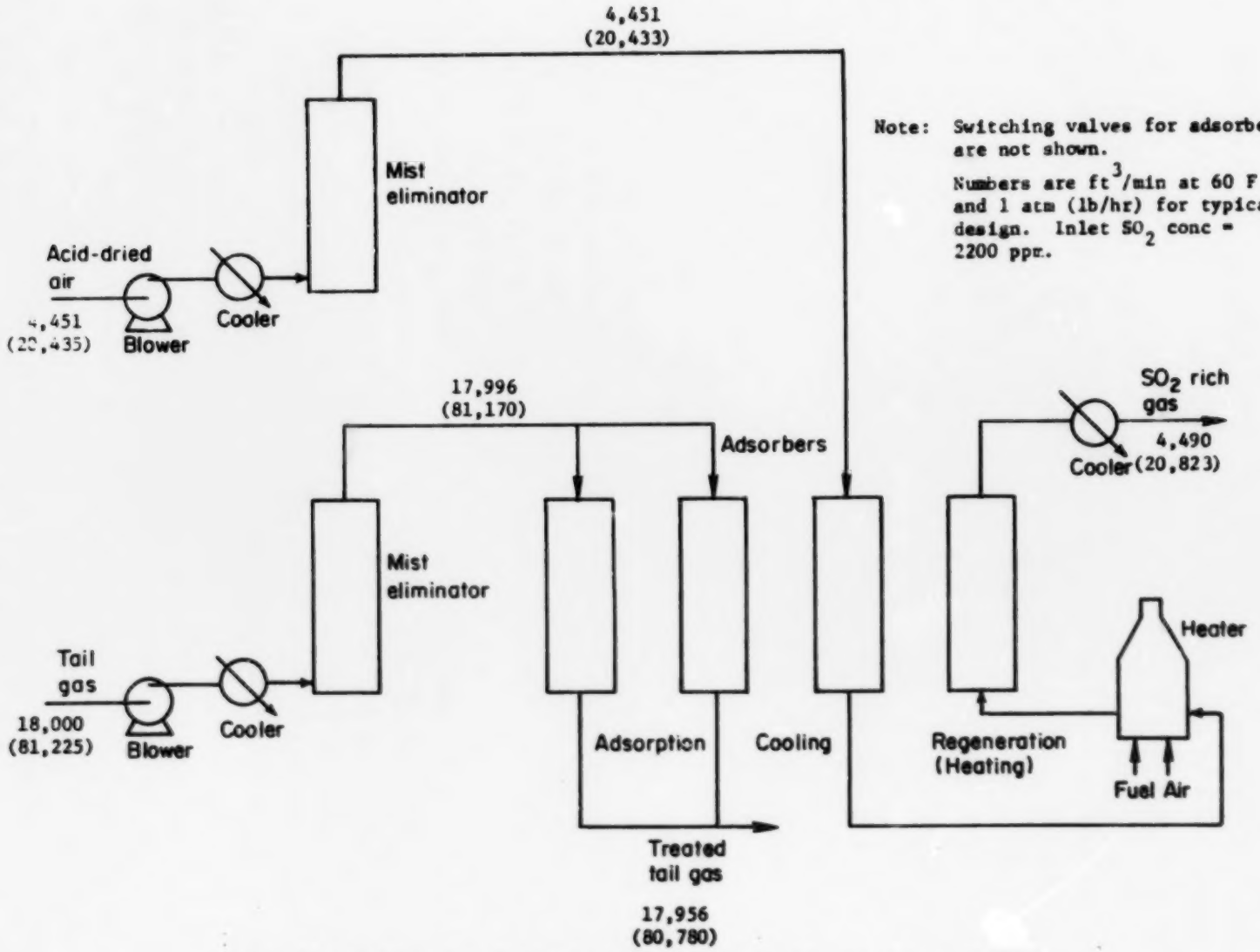


FIGURE 4. FLOW SHEET OF PURASIV S PROCESS WITH ACID-DRIED AIR REGENERATION

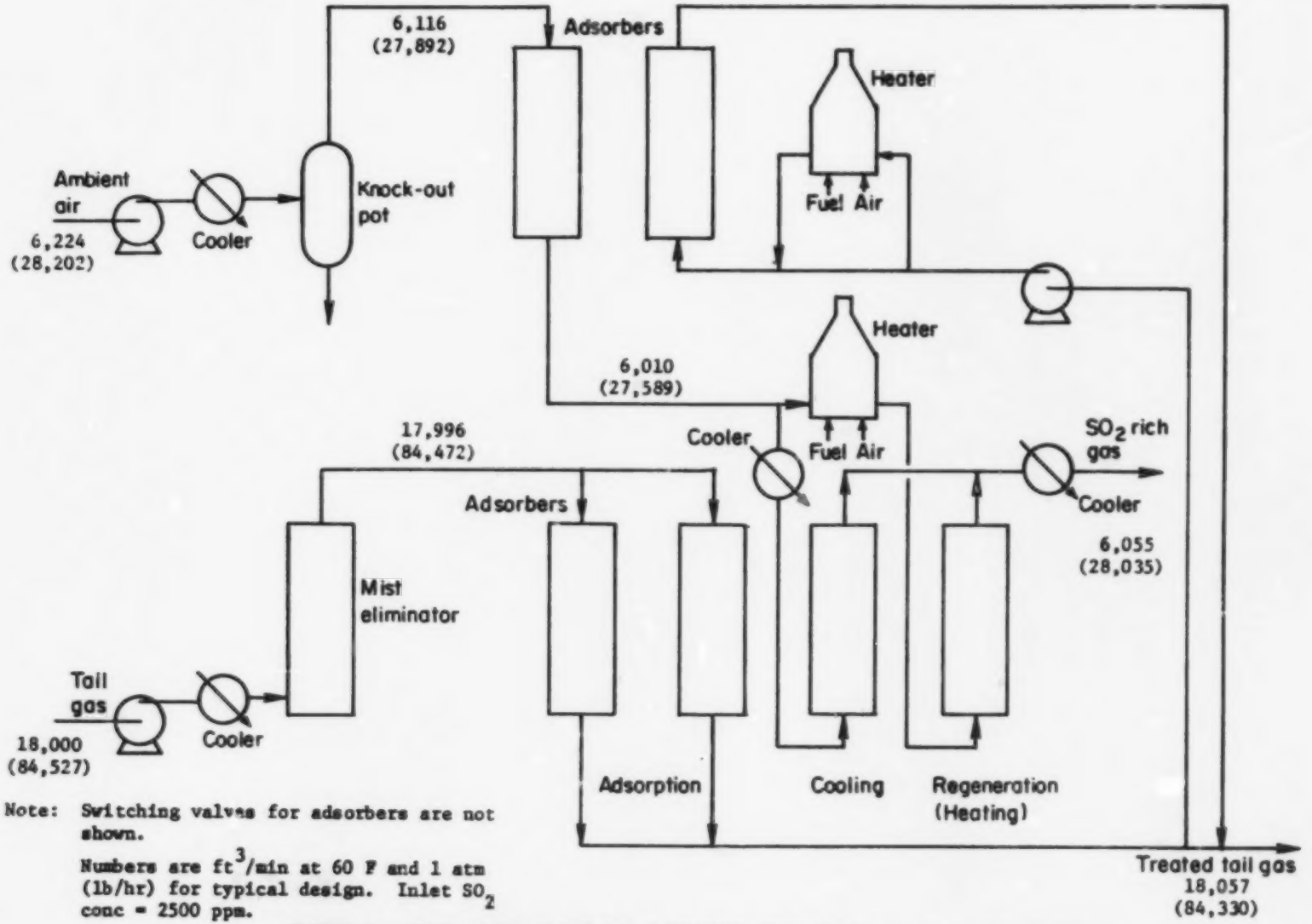
sheet for this version is shown in Figure 5. Normally, a source of acid-dried air will be available in "dry gas" sulfuric acid plants but not in "wet gas" plants.

In both versions the tail gas passes through a blower, a cooler, and a high efficiency mist eliminator which decreases the mist concentration to less than 0.1 mg/scf. The mist eliminator operates in such a way that one stage of H_2SO_4 vapor absorption by the acid wetted elements takes place. The tail gas then passes through a molecular sieve bed for selective adsorption of SO_2. The molecular sieve provides the most effective performance and longest life when the tail gas is bone dry. This is accomplished by drying the gas with a dessicant grade molecular sieve located in the same adsorber vessel in a compound bed arrangement. The treated tail gas is discharged to the atmosphere.

In the acid-dried regeneration version (Figure 4), air from the acid plant drying tower, which usually contains 50-100 ppm of water, passes through a blower, a cooler, and a high efficiency mist eliminator. This cooled regeneration gas is first used to cool the molecular sieve bed which is hot from a previous regeneration step. This heats the regeneration gas and reduces its water content to <1 ppmv. This regeneration gas then passes through a furnace where make-up heat is added and then is used to regenerate the loaded molecular sieve bed. The hot gas strips the SO_2 and water from the adsorbent. The resulting SO_2-rich gas is cooled and then recycled to the acid plant.

The four adsorber vessels shown in Figure 4 are provided with the necessary switching valves for the cyclic operation. At any given time, two adsorbers are on the adsorption step, one is on the cooling step, and one is on the regeneration (heating) step.

In the sieve-dried regeneration version (Figure 5), an additional pair of adsorbers is used to provide the dry air for regeneration. Ambient air passes through a blower, a cooler, a knock out pot, and then an adsorber containing a molecular sieve dessicant. This reduces the water content of the air to <1 ppmv. This dry air is then used to regenerate two adsorbers which are operating on a staggered cycle. Half of the regeneration gas is heated and passes through one of the adsorbers until a portion of the bed



Note: Switching valves for adsorbers are not shown.

Numbers are ft^3/min at 60 F and 1 atm (lb/hr) for typical design. Inlet SO_2 conc = 2500 ppm.

FIGURE 5. FLOW SHEET OF PURASIV S PROCESS WITH SIEVE-DRIED AIR REGENERATION

is at the regeneration temperature. This regeneration gas is then replaced with the other half of the regeneration gas, this being the portion which was not heated but rather was cooled. This cool gas cools the bed behind the hot zone and in effect "pushes" the hot zone through the bed. This is referred to as a "thermal pulse" regeneration. When the hot zone "goes out of the end" of the adsorber the bed is cool and ready for the next adsorption step. The effluent streams from the two adsorbers are combined, cooled, and recycled to the acid plant. This method of regenerating two adsorbers in parallel on a staggered cycle results in a more efficient utilization of the regeneration gas heater for the sieve-dried regeneration system.

The adsorber which was used to dry the air is periodically regenerated by the "thermal pulse" technique. A portion of the treated tail gas is used for this regeneration, and the desorbed water passes directly to the stack with the treated tail gas.

The Coulton Sludge Acid Plant

A PuraSiv S unit has been installed on one of two contact sulfuric acid plants (the "B" plant) at the Oregon, Ohio, plant of Coulton Chemical Corporation. This plant processes a mixture of alkylation spent acid (sludge) and refinery H_2S off gas. The plant is designed for a peak production rate of 200 tons per day of 99 percent H_2SO_4. The fuel used is oil.

The design basis for the Coulton PuraSiv S unit is given in Table 4. The unit was designed to reduce the SO_2 concentration of 9,450 scfm of tail gas from 3500 to less than 100 ppm. The unit includes a Brinks mist eliminator, two adsorbers for SO_2 adsorption/desorption, and two adsorbers for drying the regeneration air. Pictures of the major equipment items are shown in Figures 6, 7, and 8.

The Coulton unit went on stream February 4, 1973. The availability of the unit since that time has been very good. Immediately following startup Union Carbide performed a 6-month emission study on the system. This study showed an average SO_2 collection efficiency of 99.15 percent. Additional data from this study are discussed in the next section.

TABLE 4. DESIGN BASIS FOR PURASIV S UNIT INSTALLED AT COULTON CHEMICAL COMPANY

	Normal Flow	High Flow
Battery Limits Tail Gas Feed Conditions:		
Flow rate, (60 F, 14.7 psia) maximum	9,450 scfm	10,446 scfm
Temperature, maximum	170 F	170 F
Pressure	17.1 psia	17.5 psia
Composition, maximum concentration by volume		
SO_2	3,500 ppm	2,800 ppm
SO_3	10 ppm	10 ppm
NO_x	10 ppm	10 ppm
H_2O	100 ppm	100 ppm
H_2SO_4 (vapor + mist)	200 ppm	200 ppm
CO_2	7.5 percent	7.5 percent
O_2	10.0 percent	10.0 percent
Other minerals acids (HCl, HF, etc.)	0.2 ppm	0.2 ppm
Organic sulfides (H_2S, CS_2, etc.)	0.5 ppm	0.5 ppm
Hydrocarbons	1.0 ppm	1.0 ppm
Halogens (F_2, Cl_2, etc.)	0.5 ppm	0.5 ppm
N_2	Balance	Balance

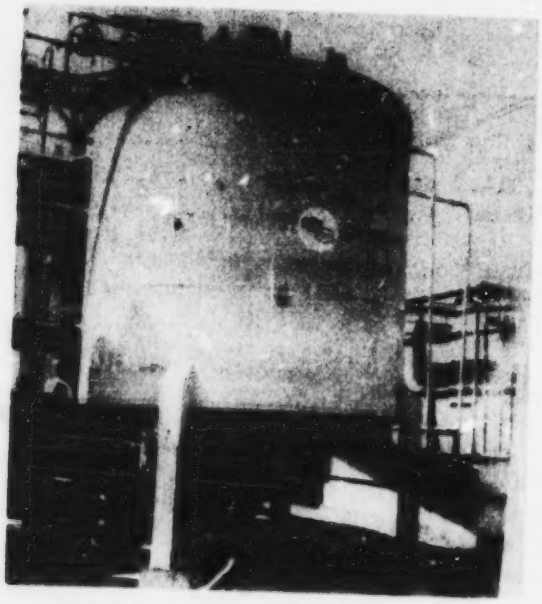


FIGURE 6. SO_2 ADSORBER

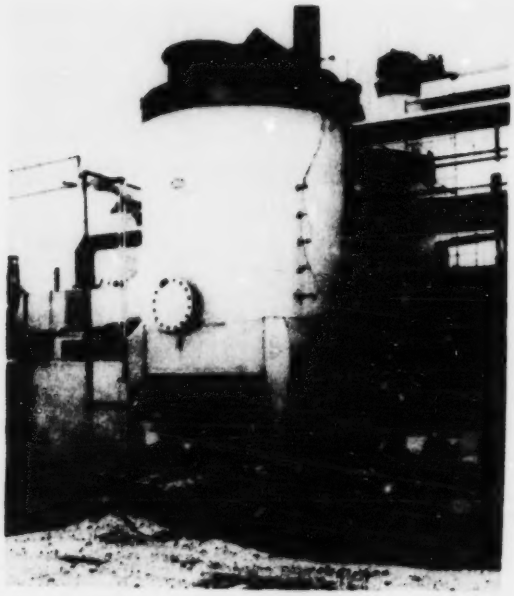


FIGURE 7. AIR DRYER

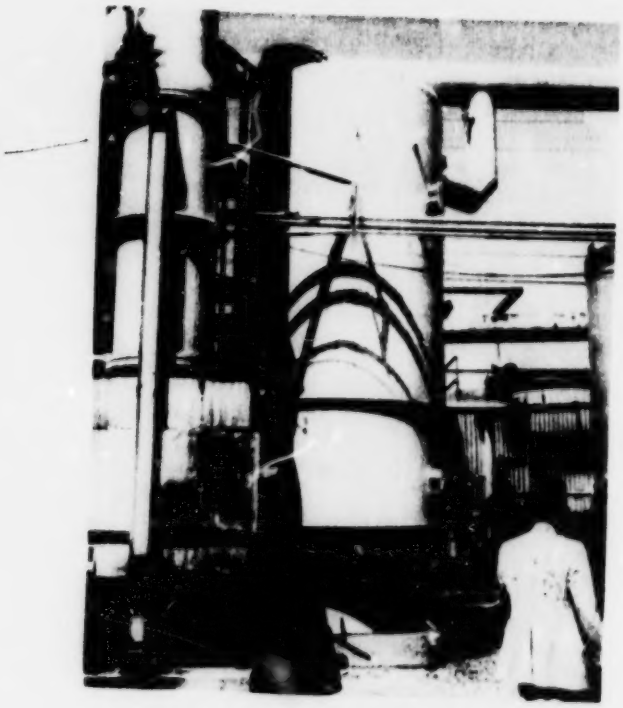


FIGURE 8. BRINK'S MIST ELIMINATOR

In November of 1973 the molecular sieve in the Coulton unit was replaced with a newer, more durable, second-generation sieve material. In February of 1975, after this sieve had seen approximately 10 months of equivalent usage, a test program was conducted by the York Research Corporation. This study showed an average SO_2 collection efficiency of 98.15 percent. The results of this test program are further discussed in a future section.

Tests by Union Carbide Corporation

The following data have been reported by Union Carbide[8] for a portion of their test program. About 2 months after startup, an extensive field evaluation of the Coulton unit was carried out by Union Carbide over a continuous period of 10 days. During this period, the tail gas flow rate varied from 8,320 to 9,860 scfm and the inlet SO_2 concentration from 2500 to 5000 ppm. In spite of these variations the unit provided an average daily effluent SO_2 concentration of 15-25 ppm. This is equivalent to an emission rate of less than 0.4 pounds of SO_2 per ton of acid and to an SO_2 recovery rate of over 99 percent.

Union Carbide has guaranteed a maximum average effluent concentration of 100 ppm SO_2 for this unit. Based on the design gas flow rate (9,450 scfm), this corresponds to a maximum emission rate of about 1.2 pounds of SO_2 per ton of acid. Thus, the above data indicate that the Coulton unit met the vendor's performance guarantee during these tests.

Tests by York Research Corporation

Sulfur Dioxide

During the 4-week test program conducted by York Research Corporation, the SO_2 concentrations in the inlet and outlet ducts of the unit were monitored with a Du Pont photometric analyzer. A total of 118 complete cycles of sieve operation were recorded during which no malfunctions or upsets occurred.

In order to determine the accuracy of the Du Pont analyzer, exit gas samples were periodically extracted and analyzed on-site using EPA Method 6. The results of these comparisons, which are discussed in the Appendix, showed that the accuracy of the Du Pont analyzer was within the limits specified by the EPA for SO_2 emissions testing.

The complete results of this test program are given in the report prepared by York Research Corporation.[9] These results will only be summarized and analyzed here.

Table 5 summarizes the results of the SO_2 emissions tests. The results are given separately for the two adsorber vessels and for three inlet SO_2 concentration ranges. With respect to the latter, "design conditions" refers to the range of 4000-5000 ppm SO_2.

The results show that adsorber A1 was operating at a somewhat lower efficiency than adsorber A2, this being due to a minor difference in design. The average emissions were 0.97 lb SO_2/ton of acid for A1 and 0.78 lb SO_2/ton of acid for A2. Both values are considerably less than the New Source Performance Standard of 4 lb SO_2/ton of acid.

The results also show that the emissions increase with increasing inlet SO_2 concentration. For adsorber A2, the emission was 0.67 lb SO_2/ton of acid for <4000 ppm SO_2 inlet, 0.87 lb/ton for 4000-4500 ppm, and 1.04 lb/ton for >4500 ppm.

The time span of this study was too short to detect any changes in the emissions with time. York recommended that a series of studies at periodic intervals be conducted for this purpose, possibly when a new molecular sieve system is installed on the "A" plant at Coulton. For a new unit such as this, York recommended a 2-month monitoring program just after startup and succeeding 1-month programs at 6-month intervals.

TABLE 5. AVERAGE SO_2 EMISSIONS - YORK RESEARCH TESTS

	Adsorber A1	Adsorber A2	Both Adsorbers
Number of Cycles Measured			
At design conditions	18	16	34 Total
At low inlet concentrations	35	34	69 Total
At high inlet concentrations	6	9	15 Total
TOTAL	59	59	118 Total
Average Emission ($lb/SO_2/T\ H_2SO_4$)			
At design conditions	1.13	0.87	1.00 Average
At low inlet concentrations	0.81	0.67	0.74 Average
At high inlet concentrations	1.36	1.04	1.20 Average
WEIGHTED AVERAGE	0.97	0.78	0.87 Average
Average Emission Rate ($lb\ SO_2/hr$)			
At design conditions	7.55	5.80	6.68 Average
At low inlet concentrations	5.07	4.20	4.64 Average
At high inlet concentrations	9.09	6.92	8.00 Average
WEIGHTED AVERAGE	6.24	5.05	5.64 Average

Other Pollutants

In addition to the SO_2 monitoring program, a series of wet chemical analyses was performed for sulfuric acid mist, total acid, chloride, sulfide, hydrocarbons, nitrogen oxides, moisture, carbon dioxide, and oxygen. The results of the acid mist analyses are summarized in Table 6 and those of other analyses in Table 7.

The emissions of sulfuric acid mist were less than 0.025 lb/ton of acid, which is far less than the New Source Performance Standard of 0.15 lb/ton of acid.

The outlet concentration was less than the inlet concentration for acid mist, chloride, hydrocarbons, and nitrogen oxides, indicating that the molecular sieve adsorbed these species to some extent. No samples of the regeneration off gas were taken, so the extent to which these species were desorbed from the sieve is unknown.

EMISSION CONTROL ALTERNATIVES

In the next section, the economic feasibility of molecular sieve emission control technology for sulfuric acid plants will be assessed by comparing this technology with other commercially available control alternatives. The two processes selected for this comparison are the Wellman-Lord Process and the dual absorption process. In this section these two processes will be described and other possible emission control alternatives will be briefly discussed.

Wellman-Lord Process

The Wellman-Lord Process is a regenerative SO_2 removal process which is available from Davy Powergas, Inc. A flow sheet for this process is shown in Figure 9. In this process the tail gas is scrubbed with an aqueous slurry containing mainly sodium sulfite (Na_2SO_3) but also some sodium bisulfite ($NaHSO_3$), sodium sulfate (Na_2SO_4), sodium hydroxide (NaOH), and sodium thiosulfate ($Na_2S_2O_3$). The absorber contains two to five

TABLE 6. SULFURIC ACID MIST EMISSIONS - YORK RESEARCH TESTS

Date	Time		Mist Concentration (mg/scf)		Mist Emission (lb/T H_2SO_4)		Mist Emission Rate (lb/hr)		Percent Emission Reduction
	Inlet	Outlet	Inlet	Outlet	Inlet	Outlet	Inlet	Outlet	
2/26	1117	1120	0.167	0.121	0.025	0.018	0.168	0.121	28
2/26	1315	1315	0.206	0.141	0.032	0.021	0.215	0.141	34
2/27	1032	1030	0.158	0.099	0.024	0.015	0.159	0.099	38
Average for 2/20 through 2/27[a]			0.215	0.071	0.034	0.011	0.216	0.071	67

(a) Five inlet measurements and six outlet measurements.

TABLE 7. OTHER EMISSIONS - YORK RESEARCH TESTS

Specie	Dates	Average Concentration (mg/scf)[a]		Average Emission Rate (lb/hr)[a]	
		Inlet	Outlet	Inlet	Outlet
Total acid (as H_2SO_4)	2/6 - 2/25	5.44	0.97	7.0[b]	1.3[b]
Sulfide (as CS_2)	2/14 - 2/21	0.53	0.59		
Chloride (as Cl^-)	2/6 - 2/25	0.1128	0.0015	0.146	0.002
Hydrocarbons (as hexane)	2/7			4.14	2.85
NO_x	2/6 & 2/25	0.00095	0.00075	0.90	0.71
Moisture	2/12 - 2/21	0.385	0.089		

(a) Each value is the average of 6 to 11 measurements.

(b) Estimated from gas flow rate calculated from chloride data (same sampling times).

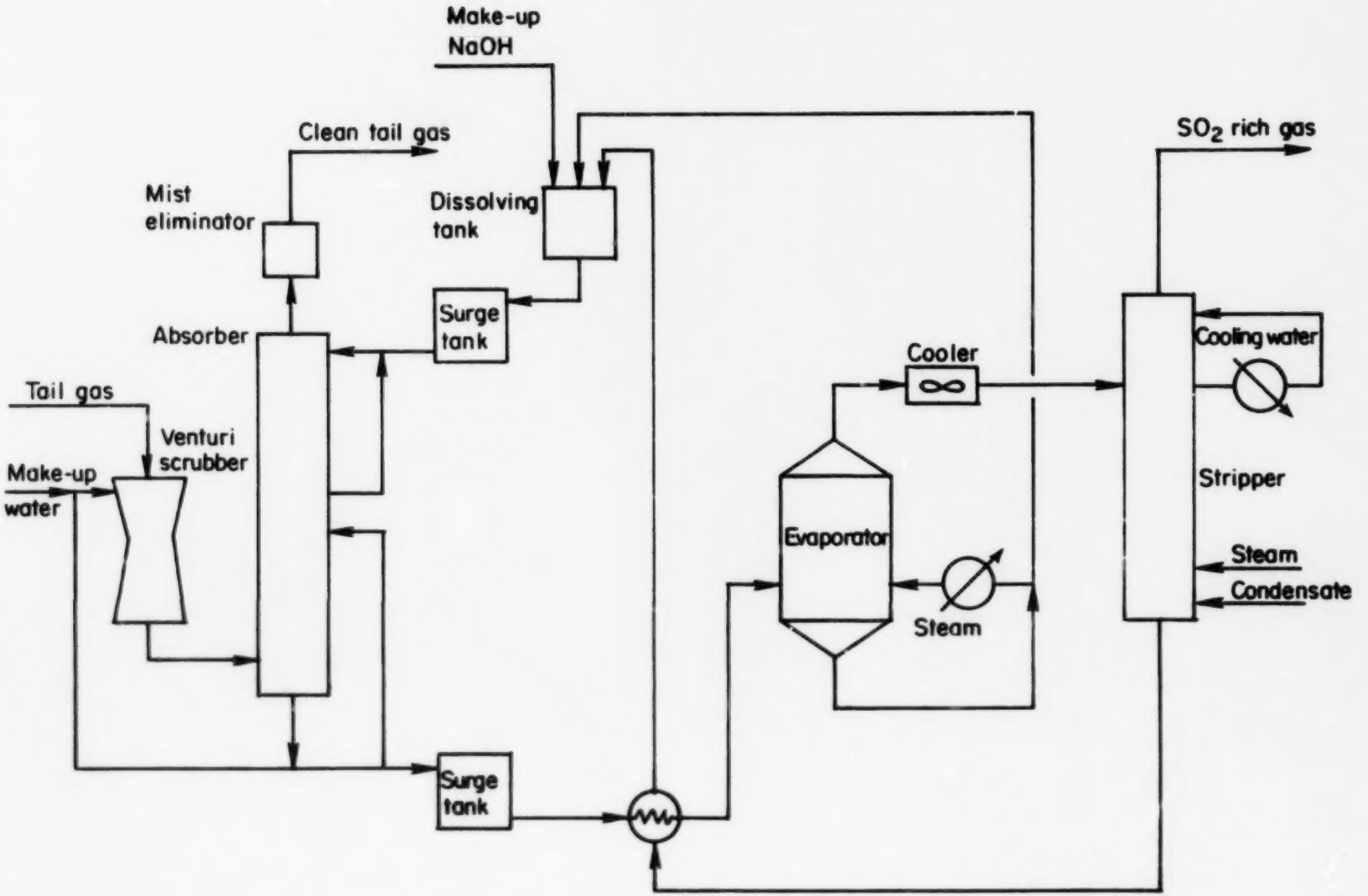


FIGURE 9. FLOW SHEET OF WELLMAN-LORD PROCESS

individually recirculated stages. The main reaction in the absorber is the conversion of the sulfite into the bisulfite as a result of SO_2 absorption:

$$SO_2(g) + Na_2SO_3(aq.) + H_2O(\ell) \rightarrow 2NaHSO_3.$$

The scrubbed gas leaves the absorber through a mist eliminator.

The liquid solution from the absorber is preheated with hot water and then goes to an evaporator, where the heat supplied causes the conversion of the bisulfite into sulfite, the latter crystallizing out because of its low solubility. The sulfite slurry from the evaporator goes to a dissolving tank where makeup caustic soda (NaOH) or soda ash (Na_2CO_3) are added. The resultant slurry is recycled to the absorber. The SO_2-rich vapor stream from the evaporator goes to a primary condenser which removes 80-90 percent of the water. The resulting condensate is steam stripped to remove the dissolved SO_2. The overhead vapors from the stripper are mixed with the vapor from the primary condenser and then pass through a secondary condenser. Additional water condensed at this point is returned to the top of the stripper. The vapor stream from the secondary condenser contains about 85 percent SO_2 and 15 percent water.

Dual Absorption Process

The dual absorption process is really an alternate design of the sulfuric acid plant which reduces the SO_2 emissions. In this process the tail gas from the primary absorber is reheated and returned to the converter where more of the SO_2 is converted to SO_3. The gas then goes to a secondary absorber where the SO_3 is absorbed to form more acid. Usually the tail gas from the primary absorber only goes through the last stage of the converter. The additional conversion of SO_2 to SO_3 is possible because the tail gas from the primary absorber has a higher O_2/SO_2 ratio than the gas originally sent to the converter (typically a ratio of >4 instead of 1.5). However, dual absorption cannot reduce the SO_2 concentration in the tail gas below about 500 ppm.

The specifics of the flow sheet for dual absorption depend on the application, the key feature being the source of heat for reheating

the tail gas from the primary absorber. For new acid plants, the dual absorption system can be integrated into the plant as shown in Figure 10.* For existing acid plants several types of add-on dual absorption systems can be used. Figure 11* shows a flow sheet for the most generally applicable version, this being one in which the necessary heat is obtained by burning additional fuel. This system is independent of the existing plant so far as heat is concerned.

For some types of acid plants it is possible to obtain at least some of the heat for the add-on dual absorption system by integration with the existing plant. One option is to install heat exchangers such that the tail gas can receive heat from the interstage coolers of the converter. Since this heat was previously used to generate steam, less steam will now be produced and the heat cost for the dual absorption system is reflected as a steam cost. Another option is to heat the tail gas by mixing in some gas bypassed from the sulfur burner. This gas will be at a temperature of 1600-1900 F and will contain 10-12 percent SO_2. Because of the additional SO_2, two conversion stages are required in this case instead of one. Also, the danger of upsets is greater for this option. Whether either of these options can be used depends on the specifics of the acid plant. They cannot be used in wet gas acid plants because either the heat is not available or the gas from the burner is too wet to be introduced into the converter.

Other Emission Control Processes

A rather thorough review of all the possible emission control systems for sulfuric acid plant tail gas has been made by Chemical Construction Corporation.[2] In this review, selections were made of the processes considered "feasible" for three different levels of SO_2 control, i.e., effluent SO_2 concentrations of <100 ppm, <250 ppm, and <500 ppm. Table 8 is an adapted listing of these selected processes which have not been completely demonstrated but are potentially feasible based on expected performance.

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* Based on Reference 2.

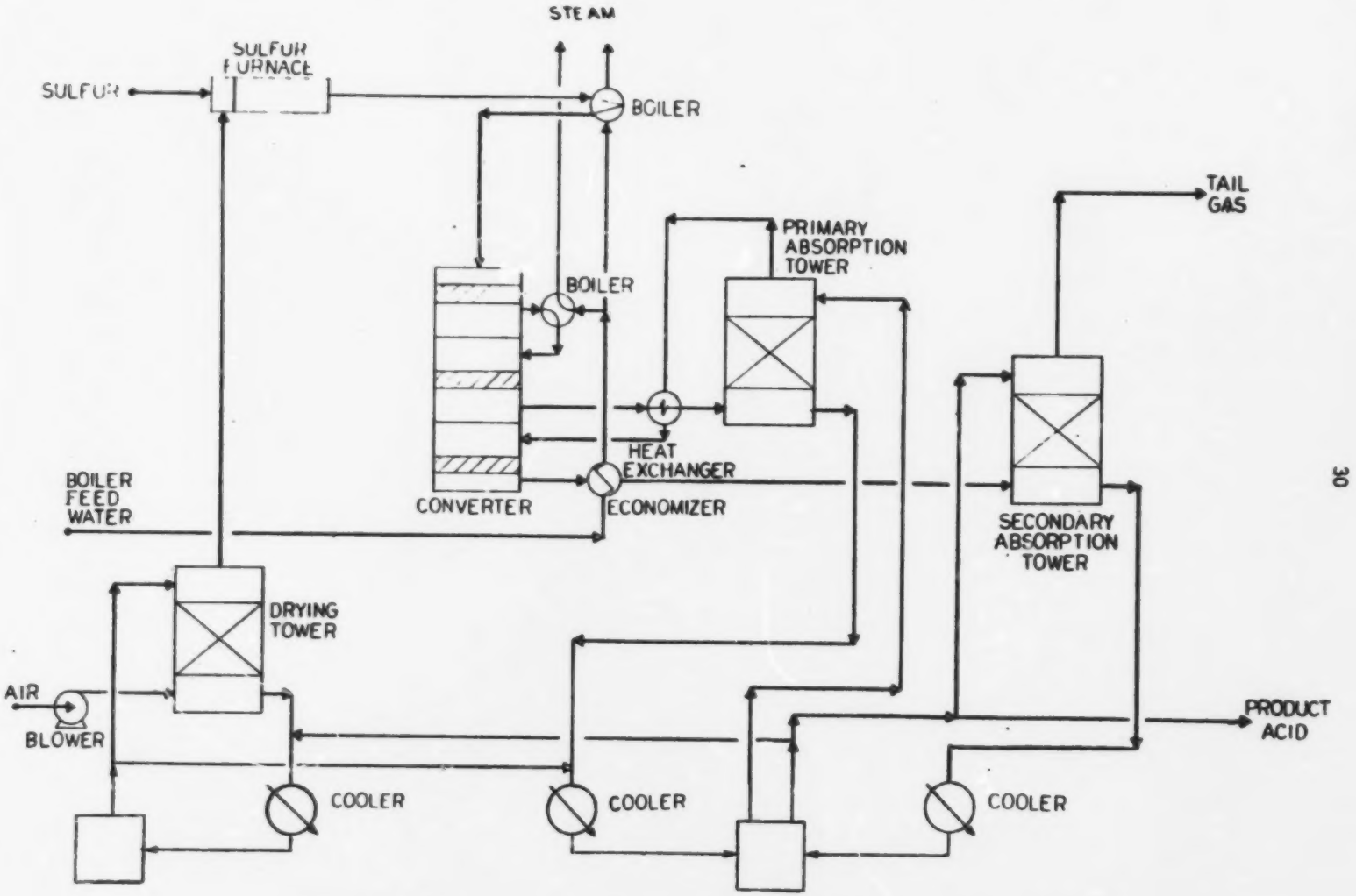

FIGURE 10. FLOW SHEET OF DUAL ABSORPTION SYSTEM FOR NEW ACID PLANTS

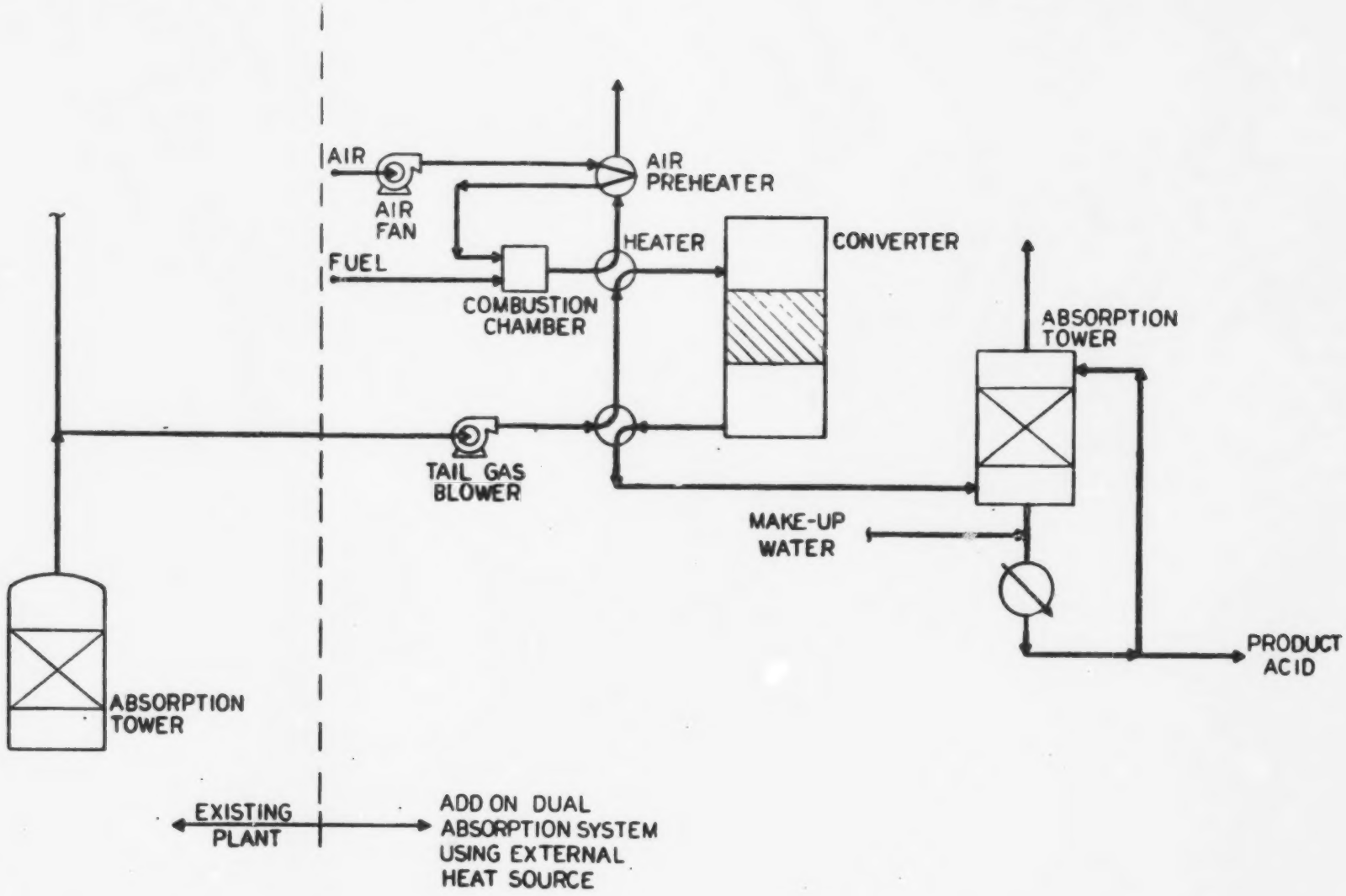

FIGURE 11. FLOW SHEET OF ADD-ON DUAL ABSORPTION PROCESS USING EXTERNAL HEAT SOURCE

TABLE 8. "FEASIBLE" EMISSION CONTROL PROCESSES FOR SULFURIC ACID PLANT TAIL GAS
Adapted from List by Chemical Construction Corp.[2]

Processes Feasible for <100 ppm SO_2 Control
• Adsorption on molecular sieves or resins
• Absorption by sodium sulfite-bisulfite system (Wellman-Lord)[a]
• Absorption by other sulfite-bisulfite systems (potassium, ammonium, methylammonium)
• Absorption by magnesium oxide
• Absorption by Na_2CO_3 to produce Na_2SO_3
• Absorption by lime
• Two-stage absorption by sulfuric acid and lime to recover SO_2 and produce plaster of Paris
Processes Feasible for 100-250 ppm SO_2 Control
• Absorption and oxidation of SO_2 in charcoal beds (Sulfacid process)
• Absorption with basic aluminum sulfate solution (Hardman-Holden)
Processes Feasible for 250-500 ppm SO_2 Control
• Dual absorption

(a) This control scheme was listed by Chemical Construction Corp., as being feasible for <100 ppm SO_2 control. However, as discussed in this report, data on the Wellman-Lord process indicate that the range of effluent concentrations is 160-210 ppm SO_2.

Scrubbing with an ammoniacal solution is a control process which has received some attention. In this process the absorption step produces ammonium bisulfite, sulfite, and sulfate. The resulting liquid is acidified with sulfuric acid and then air stripped to liberate the absorbed SO_2 for recycle to the acid plant. The resulting ammonium sulfate solution can be incorporated into a diammonium phosphate fertilizer process or can be incinerated to water vapor, nitrogen, and sulfur oxides. The major problem with this process is that, even with mist eliminators, it is very difficult to avoid a highly visible plume resulting from fine particles of ammonium sulfite and bisulfite in the scrubbed tail gas.

The three processes considered in this report are generally the most advanced processes for this application and would probably be the most attractive at this time to the operators of sulfuric acid plants. All three processes are currently being used, or at least tested, on commercial acid plants.

ECONOMIC EVALUATION OF MOLECULAR SIEVE CONTROL TECHNOLOGY

In this section the economic feasibility of molecular sieve emission control technology for sulfuric acid plants will be assessed by comparison with the two other control processes described in the previous section. This comparison will be based on the concept of capitalized costs. The following will review the basis and procedures for the cost calculations, the sources of cost data used, and the results of the calculations.

Basis and Procedures

Investments

The investments (capital costs) for the control alternatives were obtained from data on units which have been built and other data provided by system designers, vendors, and operators. The costs reported are mid-1974 estimates as updated by the Marshall and Stevens Index (= 398 in mid-1974). All calculations were made for two acid plant sizes, 200 and 1000 tons/day of 100 percent H_2SO_4. For convenience in further calculations the investments

have been expressed by an equation of the form

$$\text{Investment} = A(\text{tons/day } H_2SO_4)^B \qquad (1)$$

where A and B are constants.

The investments include not only on-site facilities but also off-site facilities associated with utilities. The off-site investments are based on the following unit costs.

Electricity	$360/kw
Cooling water	$27.50/gpm
Steam	$10.75/lb/hr

These values were determined in connection with a recent Battelle study.[10]

Operating and Maintenance Costs

The operating and maintenance (O & M) costs were calculated on an annual basis and include the following

- Operating labor
- Utilities (electricity, cooling water, steam, and fuel)
- Raw materials and chemicals
- Maintenance labor and materials
- Local property taxes and insurance
- By-product credits.

Table 9 lists the unit costs used in these calculations.

Again, the calculations were made for the two acid plant sizes. For convenience in further calculations the operating and maintenance costs have been expressed by an equation of the form

$$\text{O \& M Cost} = C(\text{tons/day } H_2SO_4)^B + D(\text{tons/day } H_2SO_4)^E \qquad (2)$$

where B, C, D, and E are constants. The first term here is the investment-related O & M costs, which includes maintenance and local taxes and insurance. Since this is taken as a percentage of the investment, the constant B is the same as the investment equation. The second term above includes all the other O & M costs.

TABLE 9. UNIT COSTS USED IN OPERATING AND MAINTENANCE COSTS

Item	Unit Cost
Electricity	1.2¢/kwhr
Cooling water	3¢/1000 gal
Process water	15¢/1000 gal
Steam	1.2$/1000 lb
Fuel oil	12$/bbl
Caustic soda (NaOH)	150$/ton
Sulfur credit	10$/ton
SO_2 credit	10$/ton
H_2SO_4 credit	25$/ton

Capitalized Costs

The economic comparison of the control alternatives was done through the concept of capitalized costs. The basic principle underlying this method of economic evaluation of alternatives is the indefinite and perpetual replacement of a given capital investment which has a finite life. The capitalized cost is defined as the original investment plus the present value of the perpetuity which will permit the perpetual replacement of the item.[11] For an item with an initial cost of I and a salvage value of S after n years the capitalized cost K (in present dollars) is

$$K = I + \frac{(I - S)}{(1 + i)^n - 1} \tag{3}$$

where i is the fractional annual interest rate or, more precisely, the rate of return on investment which the investor requires. This is usually of the order of 15 percent.[12] In this study, the capitalized cost associated with the investment, which will be designated K_I, was calculated from Equation 3 with the assumption of zero salvage value. That is,

$$K_I = \frac{I\,(1 + i)^n}{(1 + i)^n - 1}\,. \tag{4}$$

The concept of capitalized costs can be extended to include annual and other periodic costs in addition to capital investments. An expression for the capitalized cost associated with the annual O & M costs can be obtained from Equation 3 by treating this like an investment with a life of 1 year but discounting to present dollars, since these are end-of-year costs. Thus,

$$K_{O\ \&\ M} = \frac{O\ \&\ M}{i}\,. \tag{5}$$

The total capitalized cost is the sum of the term for the initial investment and that for the O & M costs.

$$K_{tot} = K_I + K_{O\ \&\ M} \tag{6}$$

In this study a 15 percent rate of return and a 12-year equipment life were used.

Investment and Operating Costs

Molecular Sieve Process

The on-site investment and operating requirements for the PuraSiv S process were based on a series of plots obtained from Union Carbide in February of 1975.* These items were plotted against the tail gas flow rate. Based on the data previously discussed, the tail gas production was taken as 85,700 scf per ton of H_2SO_4 for dry gas plants and 1.5 times that value for wet gas plants. The stated accuracy of the investment costs was ±25 percent. Other plots gave the annual cost for the adsorbent contract and the requirements of electrical power, heat, and cooling water. Separate plots were given for the acid-dried air and the sieve-dried air regeneration systems.

The O & M costs were calculated at two different levels of control, which were maximum effluent SO_2 concentrations of 100 and 500 ppm. The level of control affects only the adsorbent contract cost and the SO_2 by-product credit. The Union Carbide plots gave the adsorbent contract cost at both levels. The 100 ppm level corresponds to SO_2 emissions of about 1.5 lb SO_2/ton acid for dry gas plants or 2.3 lb SO_2/ton acid for wet gas plants. These emissions are less than the New Source Performance Standard (4 lb SO_2/ton acid). The 500 ppm level corresponds to emissions greater than the New Source Performance Standard and most state regulations but was carried through the calculations because it provides a tie-in with the dual absorption process which cannot go below the 500 ppm level. The adsorbent contract cost is the only cost item affected by the level of SO_2 control.

The operating labor requirements for the molecular sieve process are known to be very low and were taken as 1000$/yr for a 200 T/D acid plant and 2000$/yr for a 1000 T/D acid plant.

Table 10 shows the investment and O & M costs for the PuraSiv S process.

Wellman-Lord Process

The investments for the Wellman-Lord process were based on the investment for the unit installed at the sulfuric acid plant of the Olin Corporation at Curtis Bay, Maryland. Information on this installation was obtained during a recent visit to the plant by Battelle personnel.(13) This unit processes the tail gas from dry gas acid plant facilities with a

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* These plots are presented in Appendix B. It should be noted that Union Carbide claims that the costs of the PuraSiv S process have been further reduced as a result of recent development work.(16)

TABLE 10. COSTS OF PURASIV S PROCESS

8000 hours/yr Operating Time

Type of H_2SO_4 Plant	Dry Gas				Wet Gas			
Regeneration Gas Drying Agent	Acid				Molecular Sieve			
Capacity of H_2SO_4 Plant (T/D)	200		1,000		200		1,000	
Tail Gas Flow Rate (10^3 ft^3/min at 60 F and 1 atm)	12.6		63.0		18.9		94.5	
Investment (1000$, mid-1974)								
Battery limits including license fee	1,083		2,681		1,565		3,516	
Off sites								
Electricity (360$/kw)	110		551		203		1,016	
Cooling water (27.5$/gpm)	7		36		13		65	
Site preparation	1		5		1		5	
TOTAL	1,201		3,272		1,782		4,602	
Operating and Maintenance Cost (1000$/yr)(a)								
Operating labor	1.0		2.0		1.0		2.0	
Molecular sieve contract	66.8	(47.2)	334.0	(236.0)	118.5	(90.7)	592.6	(453.6)
Electricity (1.2¢/kwhr)	29.4		146.9		54.2		270.9	
Cooling water (3¢/1000 gal)	3.8		19.0		6.8		33.8	
Fuel oil (12$/bbl)	16.0		80.2		46.5		232.6	
Subtotal	117.0		582.1		227.0		1,131.9	
Maintenance (2.5 percent of inv/yr)	30.0		81.8		44.5		115.0	
Taxes and insurance (2.5 percent of inv/yr)	30.0		81.8		44.5		115.0	
Total Gross	177.0	(157.4)	745.7	(647.7)	316.0	(288.2)	1,361.9	(1,222.9)
Credit for SO_2 recovered (10$/T)	-13.3	(-11.2)	-66.3	(-56.1)	-26.0	(-22.9)	-130.0	(-114.7)
Total Net	163.7	(146.2)	679.4	(591.6)	290.0	(265.3)	1,231.9	(1,108.2)
Net Operating and Maintenance Cost in $/T Total H_2SO_4	2.38	(2.14)	1.98	(1.73)	4.11	(3.78)	3.49	(3.16)

(a) Main entry is for 100 ppm SO_2 average effluent and 2-year sieve life. Values in parentheses are for 500 ppm average effluent and 3-year sieve life.

total capacity of 1000 tons/day of H_2SO_4. The investment for the Wellman-Lord unit in mid-1972 was 2.8 million dollars, including the purge stream treating facility, site preparation, control room, instrumentation, contractor fee, etc. This investment was updated to mid-1974 using the Marshall and Stevens Index. Based on other information available on the Wellman-Lord Process,[14] half the investment was assumed to depend on the tail gas rate and half on the sulfur rate. Size exponents of 0.6 were used for scaling both portions of the investment. In this way the investments were calculated for the smaller plants and the wet gas plants.

The operating labor requirement was taken as one man per shift, based on the Olin plant. The utility and chemical requirements were based on values given by Schneider.[15] These values are as follows.

	Value for 800 T/D Acid Plant	Pertinent Ratio
Electrical power	227 kw	0.0751 kwhr/10^3 scf tail gas
Cooling water	560 gal/min	11.1 gal/10^3 scf tail gas
Steam	11,000 lb/hr	3.62 lb/10^3 scf tail gas
NaOH	2.65 tons/day	0.332 ton/ton S recovered

These requirements are in general agreement with other values given for the Wellman-Lord process,[13,14] as is shown in Appendix B.

The NaOH consumption and sulfur credit were based on 94 percent removal of SO_2 from the tail gas, since this is the maximum removal achieved by the Olin plant. Under the assumptions of this study, this removal corresponds to the following effluent SO_2 concentrations and emissions.

Type of acid plant	Dry gas	Wet gas
ppm SO_2 in effluent	162	210
lb SO_2 emitted/ton H_2SO_4	2.48	4.81

Table 11 shows the investments and O & M costs for the Wellman-Lord Process.

TABLE 11. COSTS OF WELLMAN-LORD PROCESS

8000 hours/yr Operating Time

Type of H_2SO_4 Plant	Dry Gas		Wet Gas	
Capacity of H_2SO_4 Plant (T/D)	200	1,000	200	1,000
Tail Gas Flow Rate (10^3 ft^3/min at 60 F and 1 atm)	12.6	63.0	18.9	94.5
Investment (1000$, mid-1974)				
Battery limits	1,278	3,357	1,772	4,654
Off sites				
Electricity (360$/kw)	20	102	31	153
Steam ($10.75/lb/hr)	29	147	44	221
Cooling water ($27.50	4	19	6	29
TOTAL	1,331	3,625	1,853	5,057
Operating and Maintenance Cost (1000$/yr)				
Operating labor (8$/hr)	64.0	64.0	64.0	64.0
NaOH (150$/T)[a]	32.2	161.0	62.6	312.9
Electricity (1.2¢/kwhr)	5.5	27.3	8.2	40.9
Steam (1.2$/1000 lb)	26.3	131.5	39.5	197.3
Cooling water (3¢/1000 gal)	2.0	10.1	3.0	15.1
Subtotal	130.0	393.9	177.3	630.2
Maintenance (2.5 percent of inv/yr)	33.3	90.6	46.3	126.4
Taxes and insurance (2.5 percent of inv/yr)	33.3	90.6	46.3	126.4
Total Gross	196.6	575.1	269.9	883.0
Credit for S recovered (10$/T)[a]	-6.5	-32.4	-12.6	-62.9
Total Net	190.1	542.7	257.3	820.1
Net Operating and Maintenance Cost in $/T H_2SO_4	2.85	1.63	3.86	2.46

(a) Based on 94 percent removal of SO_2 from tail gas.

Dual Absorption Process

Equations expressing the investment for the dual absorption process as a function of the acid plant capacity were developed by Battelle personnel as part of the Cost of Clean Air study.[3] For new acid plants this investment is the difference between the investment for a dual absorption plant and that for a single absorption plant. For existing plants it is the investment for the add-on system using an external heat source, although the source of heat does not greatly affect the investment. The equation used is

$$\text{Investment (\$)} = A(\text{tons/day 100 percent } H_2SO_4)^B.$$

The constant A was updated from mid-1973 to mid-1974 using the Marshall and Stevens Index. The resulting values of the constants are as follows:

Type of Acid Plant	Constant A	Constant B
Dry gas, new	14,069	0.59
Dry gas, existing	47,564	0.60
Wet gas, new	30,351	0.58
Wet gas, existing	90,017	0.61

The utility requirements for the dual absorption process were based on values given by Schneider[16] for fuel-fired units. These values are as follows.

	Value for 800 T/D Acid Plant
Electrical power	978 kw
Cooling water	580 gal/min
Fuel oil	99 gal/hr

These requirements were taken as directly proportional to the acid plant capacity.

A dual absorption acid plant requires very little operating labor beyond that for a single absorption plant. The operating labor requirements were taken as 1000$/yr for a 200 T/D plant and 2000$/yr for a 1000 T/D plant, the same values as were used for the molecular sieve process.

Table 12 shows the investments and O & M costs for the dual absorption process for new acid plants and Table 13 the corresponding costs for existing acid plants.

Summary of Costs and Equations

Table 14 summarizes the investments and operating costs for the three emission control processes. Values are given for the molecular sieve process at two levels of control (100 and 500 ppm), for two acid plant sizes (200 and 1000 tons/day), and for the two types of acid plants (dry gas and wet gas). In addition to the investments and O & M costs taken from the previous tables, Table 14 includes the total operating cost obtained by adding to the O & M cost the capital-related costs for depreciation and return on investment.

Table 15 presents the equations for the total investment and net O & M costs as functions of acid plant capacity for the molecular sieve, Wellman-Lord, and dual absorption processes.

Capitalized Costs

Table 16 shows the capitalized costs for the three emission control processes considered here. As in the previous table, values are given for the molecular sieve process at two levels of control, for two acid plant sizes, and for the two types of acid plants.

Figures 12 and 13 show the capitalized costs for dry gas and wet gas acid plants, respectively, as functions of capacity. These figures were prepared from the values in Table 16 plus values calculated for two other plant sizes (400 and 2000 tons/day).

For dry gas acid plants, the dual absorption process is the least expensive. However, it should be noted at this point that this process is also the least effective, it being limited to an effluent concentration of about 500 ppm SO_2. This concentration corresponds to emissions greater than the New Source Performance Standard and most state regulations. Of the other two processes, the molecular sieve process is less expensive for smaller acid plants (less than about 700 tons/day) and the Wellman-Lord process is less expensive for larger acid plants.

TABLE 12. COSTS OF DUAL ABSORPTION PROCESS FOR NEW ACID PLANTS

8000 hours/yr Operating Time

Type of H_2SO_4 Plant	Dry Gas		Wet Gas	
Capacity of H_2SO_4 Plant (T/D)	200	1,000	200	1,000
Tail Gas Flow Rate (10^3 ft^3/min at 60 F and 1 atm)	12.6	63.0	18.9	94.5
Investment (1000$, mid-1974)				
Battery limits	321	828	656	1,668
Off sites				
Electricity (360$/kw)	88	440	132	659
Cooling water (27.5$/gpm)	4	20	6	30
TOTAL	413	1,288	794	2,357
Operating and Maintenance Cost (1000$/yr)				
Operating labor	1.0	2.0	1.0	2.0
Electricity (1.2¢/kwhr)	23.4	117.2	35.2	175.8
Cooling water (3¢/1000 gal)	2.1	10.4	3.2	15.6
Fuel oil (12$/bbl)	56.7	283.3	85.0	425.0
Subtotal	83.2	412.9	124.4	618.4
Maintenance (4.0 percent of inv/yr)	16.5	51.5	31.8	94.3
Taxes and insurance (2.5 percent of inv/yr)	10.3	32.2	19.9	58.9
Total Gross	110.0	496.6	176.1	771.6
Credit for additional H_2SO_4 (25$/T)[a]	-42.9	-214.6	-87.8	-439.0
Total Net	67.1	282.0	88.3	332.6
Net Operating and Maintenance Cost in $/T Total H_2SO_4	0.98	0.82	1.26	0.95

(a) Based on 500 ppm SO_2 in treated tail gas.

TABLE 13. COSTS OF ADD-ON DUAL ABSORPTION PROCESS FOR EXISTING ACID PLANTS

8000 hours/yr Operating Time

Type of H_2SO_4 Plant	Dry Gas		Wet Gas	
Capacity of H_2SO_4 Plant (T/D)	200	1,000	200	1,000
Tail Gas Flow Rate (10^3 ft^3/min at 60 F and 1 atm)	12.6	63.0	18.9	94.5
Investment (1000$, mid-1974)				
Battery limits	1,143	3,001	2,280	6,086
Off sites				
Electricity (360$/kw)	88	440	132	659
Cooling water (27.5$/gpm)	4	20	6	30
TOTAL	1,235	3,461	2,418	6,775
Operating and Maintenance Cost (1000$/yr)				
Operating labor	1.0	2.0	1.0	2.0
Electricity (1.2¢/kwhr)	23.4	117.2	35.2	175.8
Cooling water (3¢/1000 gal)	2.1	10.4	3.2	15.6
Fuel oil (12$/bbl)	56.7	283.3	85.0	425.0
Subtotal	83.2	412.9	124.4	618.4
Maintenance (4.0 percent of inv/yr)	49.4	138.4	96.7	271.0
Taxes and insurance (2.5 percent of inv/yr)	30.9	86.5	60.5	169.4
Total Gross	163.5	637.8	281.6	1,058.8
Credit for additional H_2SO_4 (25$/T)[a]	-42.9	-214.6	-87.8	-439.0
Total Net	120.6	423.2	193.8	619.8
Net Operating and Maintenance Cost in $/T Total H_2SO_4	1.76	1.24	2.76	1.77

(a) Based on 500 ppm SO_2 in treated tail gas.

TABLE 14. INVESTMENT AND OPERATING COSTS FOR EXISTING ACID PLANTS

Control Process		Molecular Sieve	Molecular Sieve	Wellman-Lord	Dual Absorption
ppm SO_2 in Effluent		100	500	160-210	500
Size and Type of Plant	Cost Item[a]				
200 T/D dry gas	Investment (M$)	1,201	1,201	1,331	1,235
	Operating cost (M$/yr)				
	O & M	163.7	146.2	190.1	120.6
	Capital[b]	221.6	221.6	245.5	227.8
	TOTAL	385.3	367.8	435.6	348.4
1000 T/D dry gas	Investment (M$)	3,272	3,272	3,625	3,461
	Operating cost (M$/yr)				
	O & M	679.4	591.6	542.7	423.2
	Capital[b]	603.6	603.6	668.7	638.5
	TOTAL	1,283.0	1,195.2	1,211.4	1,061.7
200 T/D wet gas	Investment (M$)	1,782	1,782	1,853	2,418
	Operating cost (M$/yr)				
	O & M	290.0	265.3	257.3	193.8
	Capital[b]	328.7	328.7	341.8	446.1
	TOTAL	618.7	594.0	599.1	639.9
1000 T/D wet gas	Investment (M$)	4,602	4,602	5,057	6,775
	Operating cost (M$/yr)				
	O & M	1,231.9	1,108.2	820.1	619.8
	Capital[b]	849.0	849.0	932.9	1,249.9
	TOTAL	2,080.9	1,957.2	1,753.0	1,869.7

(a) M = 1000.

(b) Includes return on investment at 15 percent/yr and sinking-fund depreciation for 12-year life at 15 percent/yr interest.

TABLE 15. EQUATIONS FOR INVESTMENT AND OPERATING COSTS

Control Process	Type of Acid Plant	Total Investment (1000$)[a]	Net Operating + Maintenance Cost ($/yr)	ppm SO_2 in Effluent Gas
Molecular sieve	Dry gas	$44.26\ C^{0.623}$	$526.7\ C^{0.997} + 2,213\ C^{0.623}$	100
Molecular sieve	Wet gas	$78.66\ C^{0.589}$	$1,015.8\ C^{0.998} + 3,933\ C^{0.589}$	100
Molecular sieve	Dry gas	$44.26\ C^{0.623}$	$440.1\ C^{0.996} + 2,213\ C^{0.623}$	500
Molecular sieve	Wet gas	$78.66\ C^{0.589}$	$890.7\ C^{0.998} + 3,933\ C^{0.589}$	500
Wellman-Lord	Dry gas	$49.03\ C^{0.623}$	$3,606\ C^{0.667} + 2,452\ C^{0.623}$	162
Wellman-Lord	Wet gas	$67.91\ C^{0.624}$	$2,816\ C^{0.768} + 3,396\ C^{0.624}$	210
Dual Absorption[b]	Dry gas	$41.60\ C^{0.640}$	$212.5\ C^{0.990} + 2,704\ C^{0.640}$	500
Dual Absorption[b]	Wet gas	$81.44\ C^{0.640}$	$195.0\ C^{0.988} + 5,294\ C^{0.640}$	500

(a) Includes offsites.

(b) Add-on process for existing acid plant.

TABLE 16. CAPITALIZED COSTS FOR EXISTING ACID PLANTS

Capitalized Cost (1000$)

Control Process		Molecular sieve	Molecular sieve	Wellman-Lord	Dual Absorption
ppm SO_2 in Effluent		100	500	160-210	500
Size and Type of Plant					
200 T/D dry gas	Invest.	1,477	1,477	1,637	1,519
	O&M	1,091	975	1,267	804
	TOTAL	2,568	2,452	2,904	2,323
1000 T/D dry gas	Invest.	4,025	4,025	4,458	4,257
	O&M	4,529	3,944	3,618	2,821
	TOTAL	8,554	7,969	8,076	7,078
200 T/D wet gas	Invest.	2,192	2,192	2,279	2,974
	O&M	1,933	1,769	1,715	1,292
	TOTAL	4,125	3,961	3,994	4,266
1000 T/D wet gas	Invest.	5,660	5,660	6,219	8,332
	O&M	8,213	7,388	5,467	4,132
	TOTAL	13,873	13,048	11,686	12,464

Note: Investment capitalized cost based on 12-year life; both investment and O & M capitalized costs based on 15 percent/year "interest" rate.

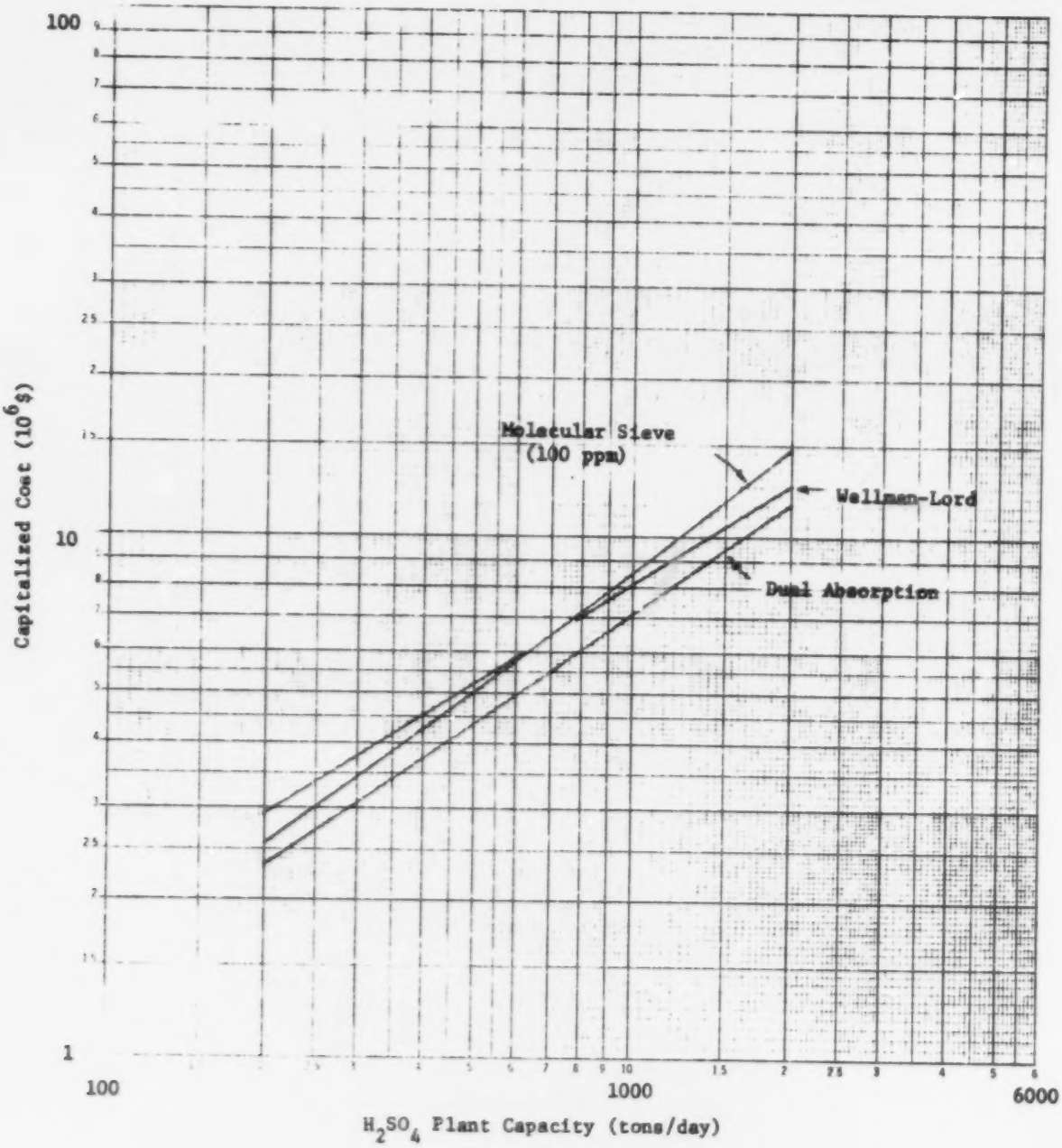

FIGURE 12. CAPITALIZED COSTS FOR SO_2 CONTROL OF DRY GAS ACID PLANTS

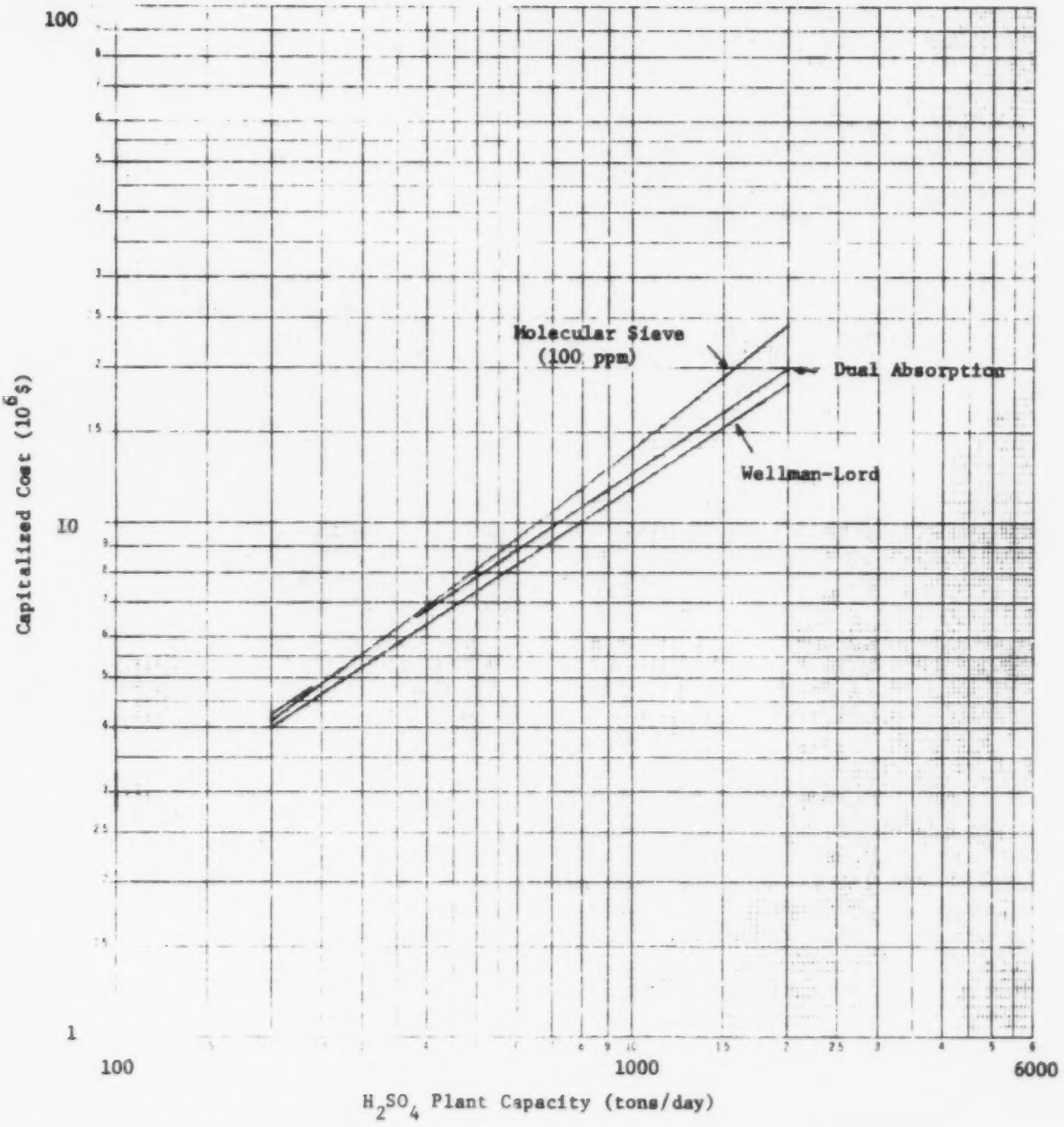

FIGURE 13. CAPITALIZED COSTS FOR SO_2 CONTROL OF WET GAS ACID PLANTS

Again, it should be noted that the molecular sieve process is capable of a higher degree of SO_2 removal than is the Wellman-Lord process (<100 ppm versus 160-210 ppm SO_2), although both can meet the New Source Performance Standard for most plants.

For wet gas acid plants, the Wellman-Lord process is the least expensive. Dual absorption has the next lowest cost except for very small acid plants, where the molecular sieve process becomes slightly less expensive.

The accuracy of the costs used here does not permit precise definition of the economic breakpoints, particularly since the costs for the three processes are quite close at all capacitier. The point is that the costs are close enough that the molecular sieve process appears to be economically competitive, at least for the smaller acid plants.

CONCLUSIONS AND RECOMMENDATIONS

This analysis of the applicability of molecular sieve technology to the control of SO_2 emissions from sulfuric acid plants leads to the following conclusions:

- Molecular sieve technology can control the SO_2 emissions to well within the most stringent existing regulation for sulfuric acid plants (4 lb SO_2/ton acid). Union Carbide field test data indicated that shortly after startup the Coulton unit was limiting the emissions to no more than 0.4 lb SO_2/ton acid. The data taken by York Research Corporation after the equivalent of 10 months of operation on the second charge of sieve showed an average emission of 0.9 lb SO_2/ton acid. Union Carbide's performance guarantee for this unit corresponds to an emission of about 1.2 lb SO_2/ton acid.
- The PuraSiv S process can limit the SO_2 emissions more effectively than either of the other processes being used or tested commercially on acid plants. The average effluent SO_2 concentration is <100 ppm, compared to 160-210 ppm for the Wellman-Lord process and about 500 ppm for the dual absorption process.
- The SO_2 emissions from a molecular sieve control unit increase with increasing inlet (tail gas) SO_2 concentration. At the Coulton unit the average emission for the cycles in which the inlet SO_2 concentration exceeded 4500 ppm was 1.2 lb SO_2/ton acid.
- The assumed 2-year sieve life with acceptable SO_2 control has not been demonstrated, although there is no reason to believe it cannot be achieved. The field tests to date have been too short to detect any change in the emissions with time.
- Molecular sieve technology appears to be economically competitive with the Wellman-Lord and dual absorption processes, at least for the smaller acid plants. The capitalized costs for all three processes are fairly close, but the dual absorption process is the least expensive for dry gas acid plants and the Wellman-Lord process is the least expensive for wet gas plants. However, the dual absorption process is limited in its effectiveness such that for most plants it cannot meet the regulation of 4 lb SO_2/ton acid (although it

can meet some less stringent regulations). Individual acid plant characteristics will affect the economic choice among these alternates.

The only recommendation arising from this study is that the 2-year sieve life be demonstrated by monitoring the SO_2 emissions at a commercial unit periodically over that time period.

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APPENDIX A

ACCURACY OF SO_2 MEASUREMENTS

This Appendix presents and expands upon an analysis by York Research Corporation of the accuracy of their SO_2 concentration measurements. These measurements were made using a Du Pont 460/1 photometric analyzer. The accuracy of this analyzer was assessed by comparing its results with the results of simultaneous measurements using EPA Method 6.

Table A-1 presents the data from the accuracy tests. Included are seven points for inlet samples with SO_2 concentrations of 2500-3400 ppm and seven points for outlet samples with SO_2 concentrations of 10-135 ppm. For each point, the absolute difference between the two analytical methods, D, and its square are tabulated.

Statistical calculations were made for three data sets - the seven inlet samples, the seven outlet samples, and all 14 samples together. For each set, the average absolute deviation was calculated as

$$\text{Average Absolute Deviation} = D_A = \frac{1}{n} \sum_{i=1}^{n} D_i$$

where n is the number of data points. The root mean square deviation was calculated as

$$\text{Root Mean Square Deviation} = \frac{1}{n}\left[\sum_{i=1}^{n} D_i^2 \right]^{1/2}.$$

The 95 percent confidence interval was calculated as

$$C_{95} = \frac{t}{n\sqrt{n-1}} \left[n \sum_{i=1}^{n} D_i^2 - \sum_{i=1}^{n} D_i^2 \right]^{1/2}$$

where t is the 2.5 percent point from the "t distribution" for n - 1 degrees of freedom. For n = 7 the value of t is 2.447 and for n = 14 it is 2.160. The accuracy of the Du Pont measurements, expressed as a percentage, was then calculated as

TABLE A-1. SO_2 MEASUREMENT ACCURACY TESTS BY YORK RESEARCH

Location	Date	Time	Test No.	vppm SO_2 by EPA Method 6	vppm SO_2 by Du Pont	Difference D	D^2
Inlet	2/12	1035	1	3322	3300	22	484
	2/12	1555	3	3359	3385	26	676
	2/14	0924	4A	3176	3217	41	1,681
	2/14	0925	4B	3022	3217	195	38,025
	2/17	1133	7	2615	2500	115	13,225
	2/17	1320	9	2749	2525	224	50,176
	2/18	1110	13	3354	3400	46	2,116
					Sum	669	106,383
Outlet	2/12	1147	2	72.3	73.	0.7	0.49
	2/14	1134	5	115.	102.	13.	169.
	2/14	1224	6	15.1	7.5	7.6	57.76
	2/17	1135	8	88.3	84.	4.3	18.49
	2/17	1340	10	16.2	10.5	5.7	32.49
	2/17	1435	11	45.7	39.	6.7	44.89
	2/18	1116	12	120.6	135.	14.4	207.36
					Sum	52.4	530.48
					Grand Sum	721.4	106,913.48

$$\text{Accuracy (\%)} = \frac{(D_A + C_{95})(100)}{\frac{1}{n}\sum_{i=1}^{n} V_i}$$

where V_i is the value (SO_2 concentration) by EPA Method 6. This definition is the one specified in the Federal Register* for SO_2 emissions testing.

The results of the statistical calculations are shown in Table A-2. The accuracy of the Du Pont analyzer is about 6 percent for the high concentration data and for the data as a whole, but about 18 percent for the low concentration data alone. Even this value is less than the 20 percent accuracy limit specified in the Federal Register.

* "Performance Specifications and Specification Test Procedures for Monitors of SO_2 and NO_x from Stationary Sources", Federal Register, 39 (177), September 11, 1974.

TABLE A-2. SO_2 MEASUREMENT ACCURACY PARAMETERS - YORK RESEARCH TESTS

Type of Samples	Inlet	Outlet	Inlet & Outlet
SO_2 Concentration Range (ppm)	2500-3400	10-135	10-3400
Number of Samples Compared	7	7	14
Comparison of Du Pont Analyzer with EPA Method 6			
Average absolute difference (ppm)	95.6	7.5	51.5
Root mean square difference (ppm)	46.6	3.3	23.4
95 percent confidence interval (ppm)	77.8	4.4	42.3
Accuracy (percent)	5.62	17.64	5.95

APPENDIX B

DETAILED COST DATA

This Appendix presents some details related to the cost data discussed in the text. The first section deals with the PuraSiv S process and the second section with the Wellman-Lord process.

PuraSiv S Process

This section includes the series of plots obtained from Union Carbide on the costs and cost-related quantities for the PuraSiv S process. The following figures are included:

Figure	Title
B-1	Investment for PuraSiv S Process with Acid-Dried Regeneration
B-2	Adsorbent Contract Cost, Acid-Dried Regeneration, Two Years, 100 ppm SO_2
B-3	Adsorbent Contract Cost, Sieve-Dried Regeneration, Two Years, 100 ppm SO_2
B-4	Adsorbent Contract Cost, Acid-Dried Regeneration, Three Years, 500 ppm SO_2
B-5	Adsorbent Contract Cost, Sieve-Dried Regeneration, Three Years, 500 ppm SO_2
B-6	Power Requirements, Acid-Dried Regeneration
B-7	Power Requirements, Sieve-Dried Regeneration
B-8	Absorbed Heat Requirements, Acid-Dried Regeneration
B-9	Absorbed Heat Requirements, Sieve-Dried Regeneration
B-10	Cooling Water Requirements, Acid-Dried Regeneration
B-11	Cooling Water Requirements, Sieve-Dried Regeneration.

These figures present the cost-related quantities used in the cost estimates presented in this report. However, it should be noted that in recent communication* Union Carbide claimed that their recent development work on the PuraSiv S process has resulted in some further

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* Cassetty, R. W., Union Carbide Corporation, Tarrytown, New York, personal communication to D. W. Hissong of BCL, August 1, 1975.

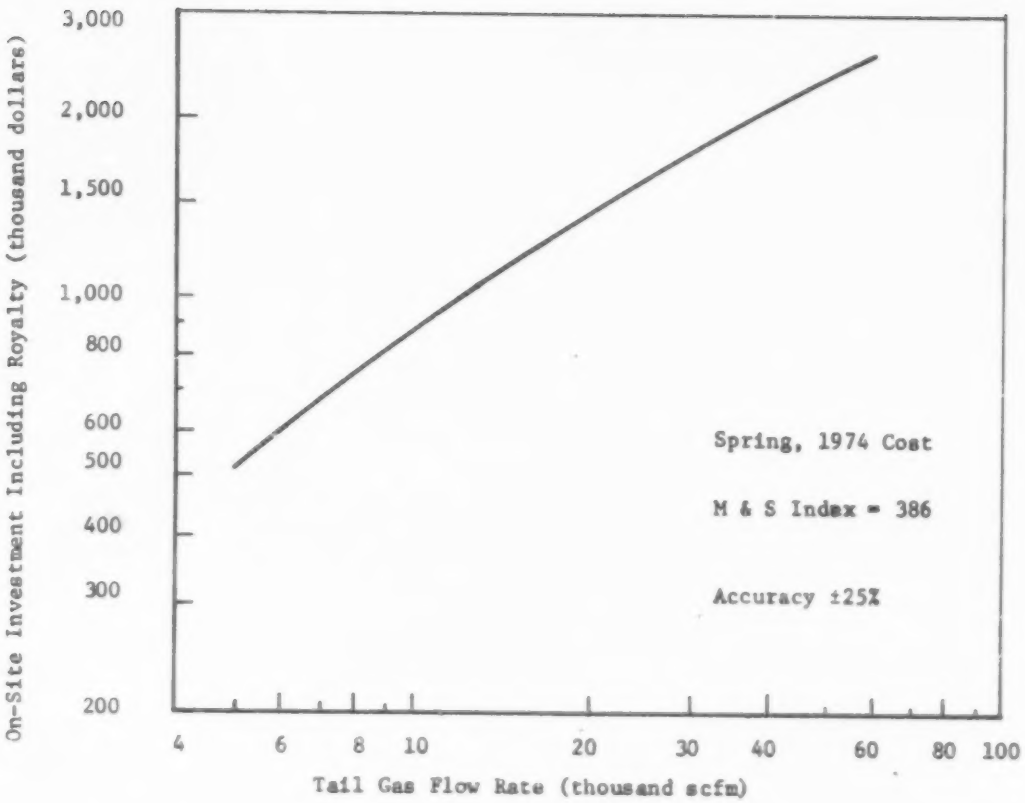

FIGURE B-1. INVESTMENT FOR PURASIV S PROCESS WITH ACID-DRIED REGENERATION

Notes:

(1) Costs shown do not include off sites, motor control center, site preparation or molecular sieve adsorbent.

(2) Costs shown are for tail gas streams containing less than 3000 ppmv SO_2. For streams containing 4000 ppmv SO_2, add 10 percent. Not applicable to tail gas streams containing more than 4000 ppmv SO_2.

(3) For a molecular sieve dried air regeneration system or for a UCC supplied acid dried air regeneration system, add 10% to above shown costs for tail gas streams containing less than 3000 ppmv SO_2. Add an additional 10% for streams containing 4000 ppmv SO_2. Not applicable to tail gas streams containing more than 4000 ppmv SO_2.

(4) In these figures, scfm = cubic feet per minute at 60 F and atmospheric pressure.

8/19/74

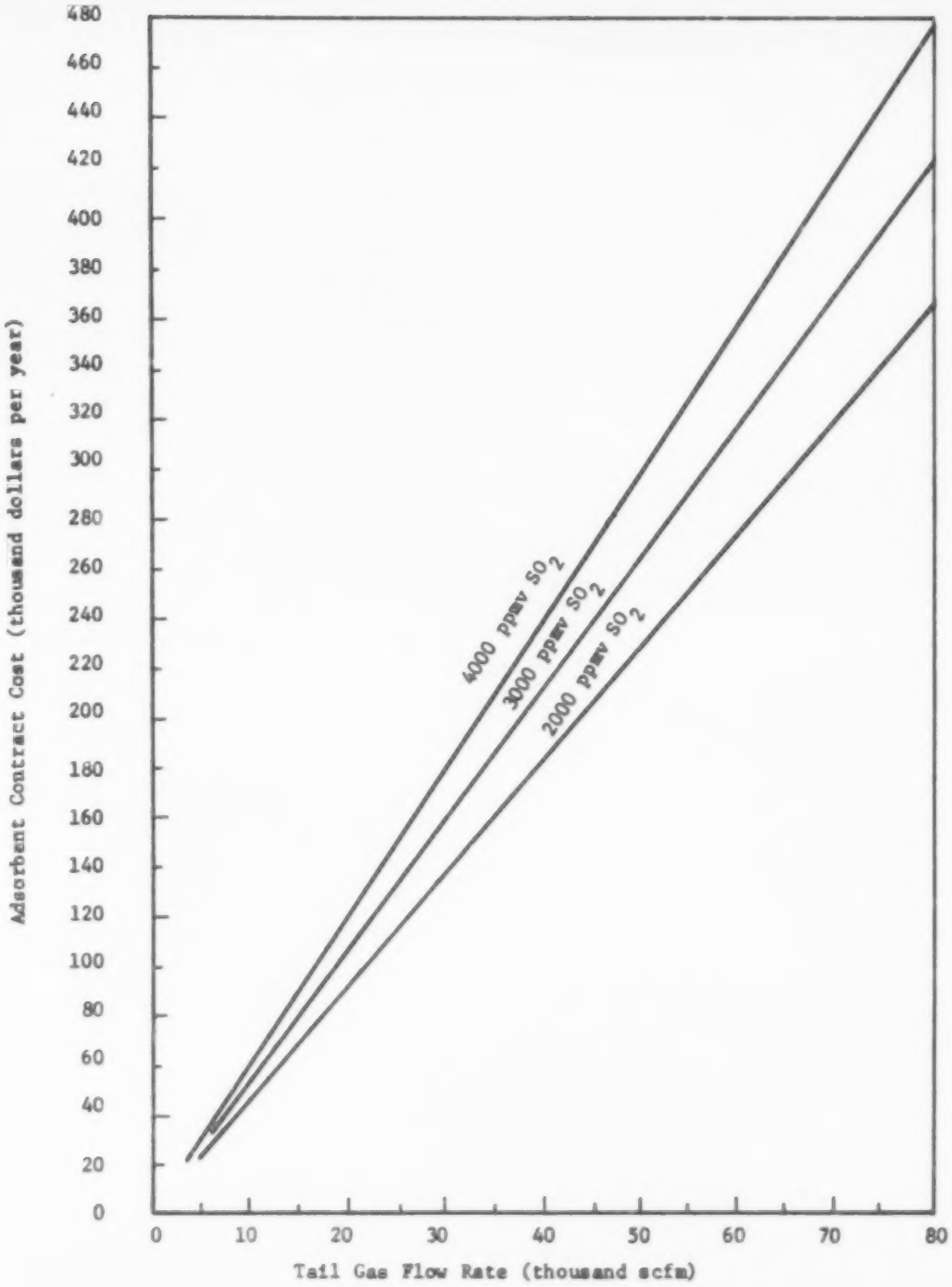

FIGURE B-2. ADSORBENT CONTRACT COST, ACID-DRIED REGENERATION, TWO YEARS, 100 PPM SO_2

12/12/74

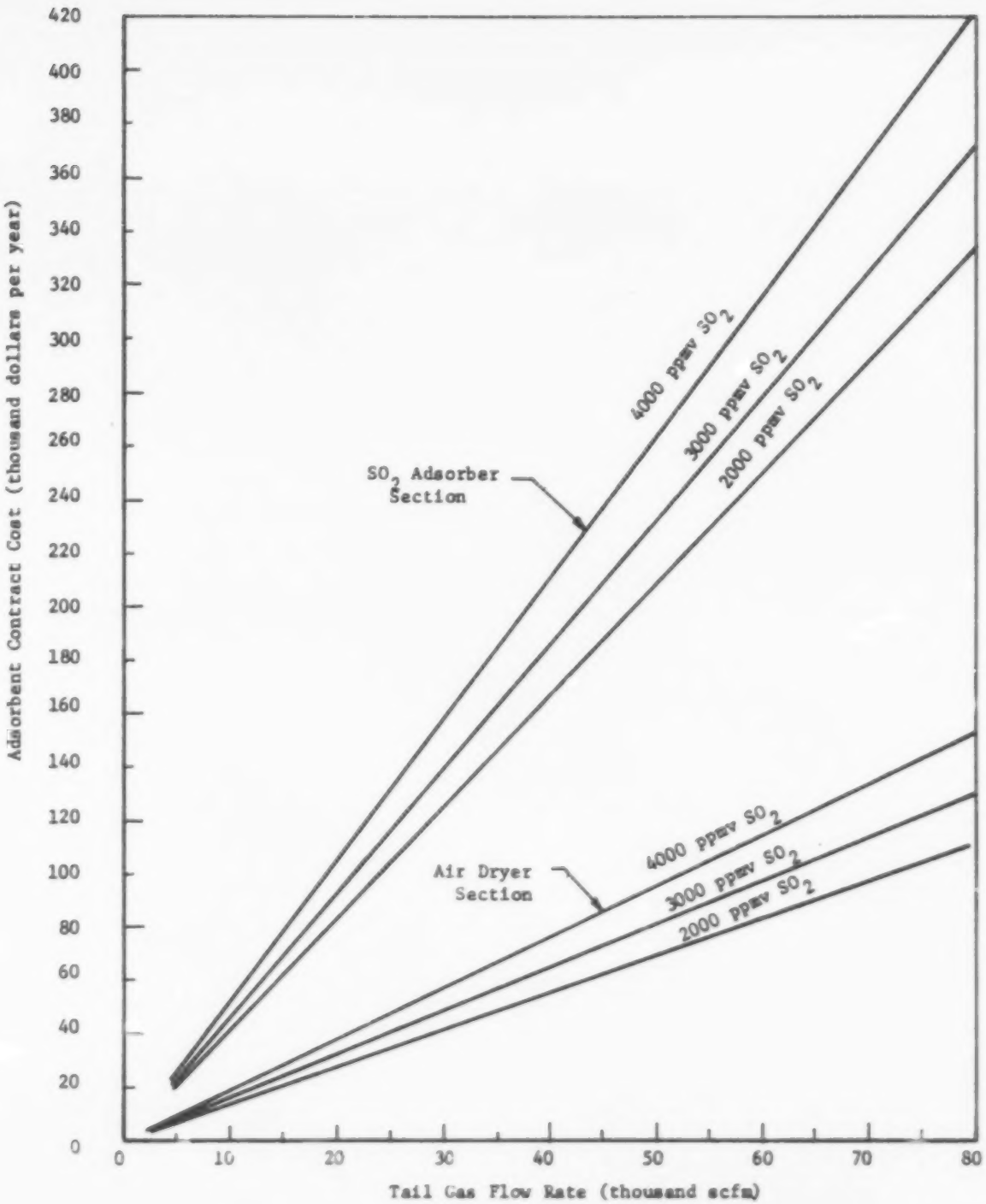

FIGURE B-3. ADSORBENT CONTRACT COST, SIEVE-DRIED REGENERATION, TWO YEARS, 100 PPM SO_2

12/12/74

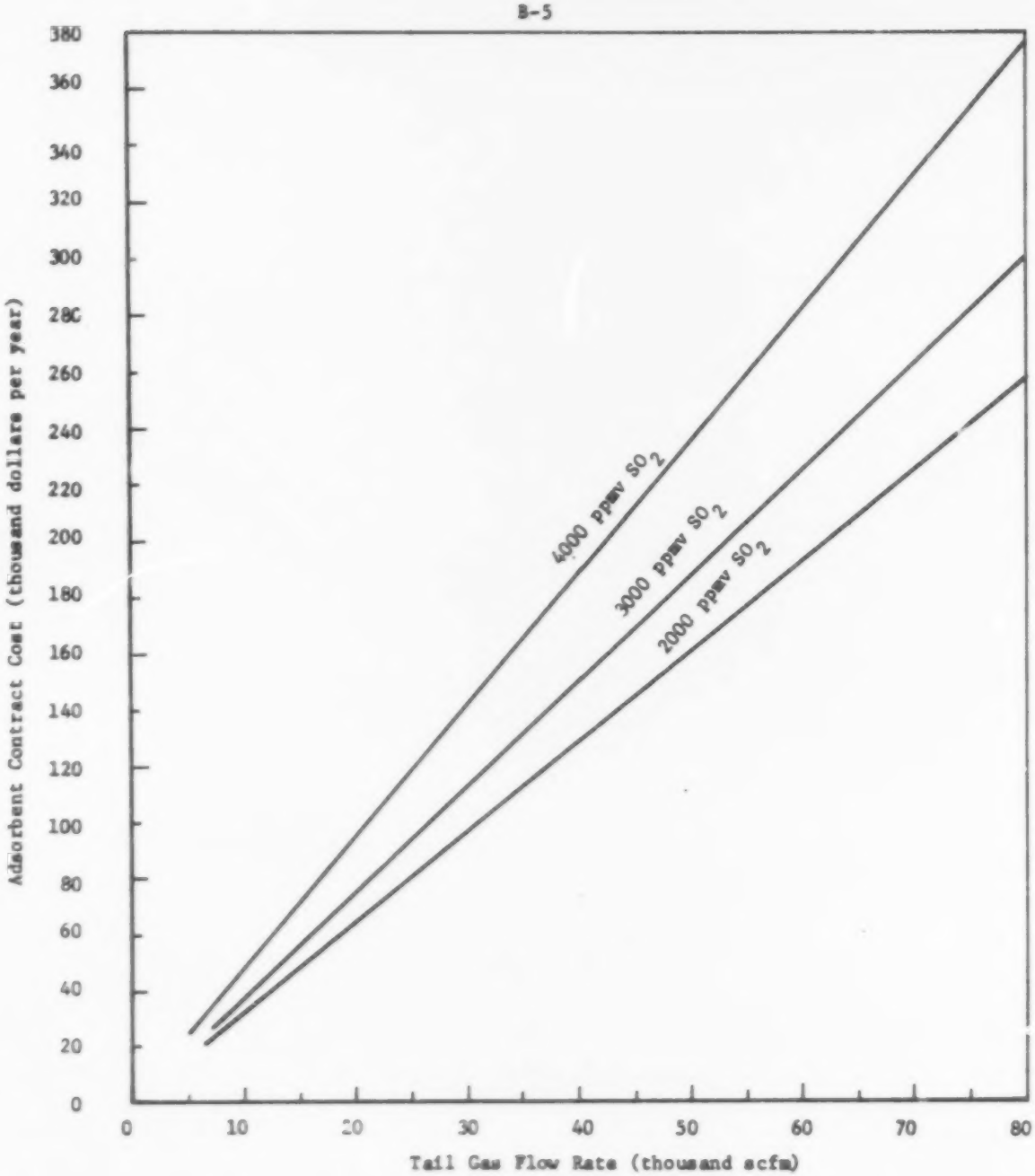

FIGURE B-4. ADSORBENT CONTRACT COST, ACID-DRIED REGENERATION, THREE YEARS, 500 PPM SO_2

12/12/74

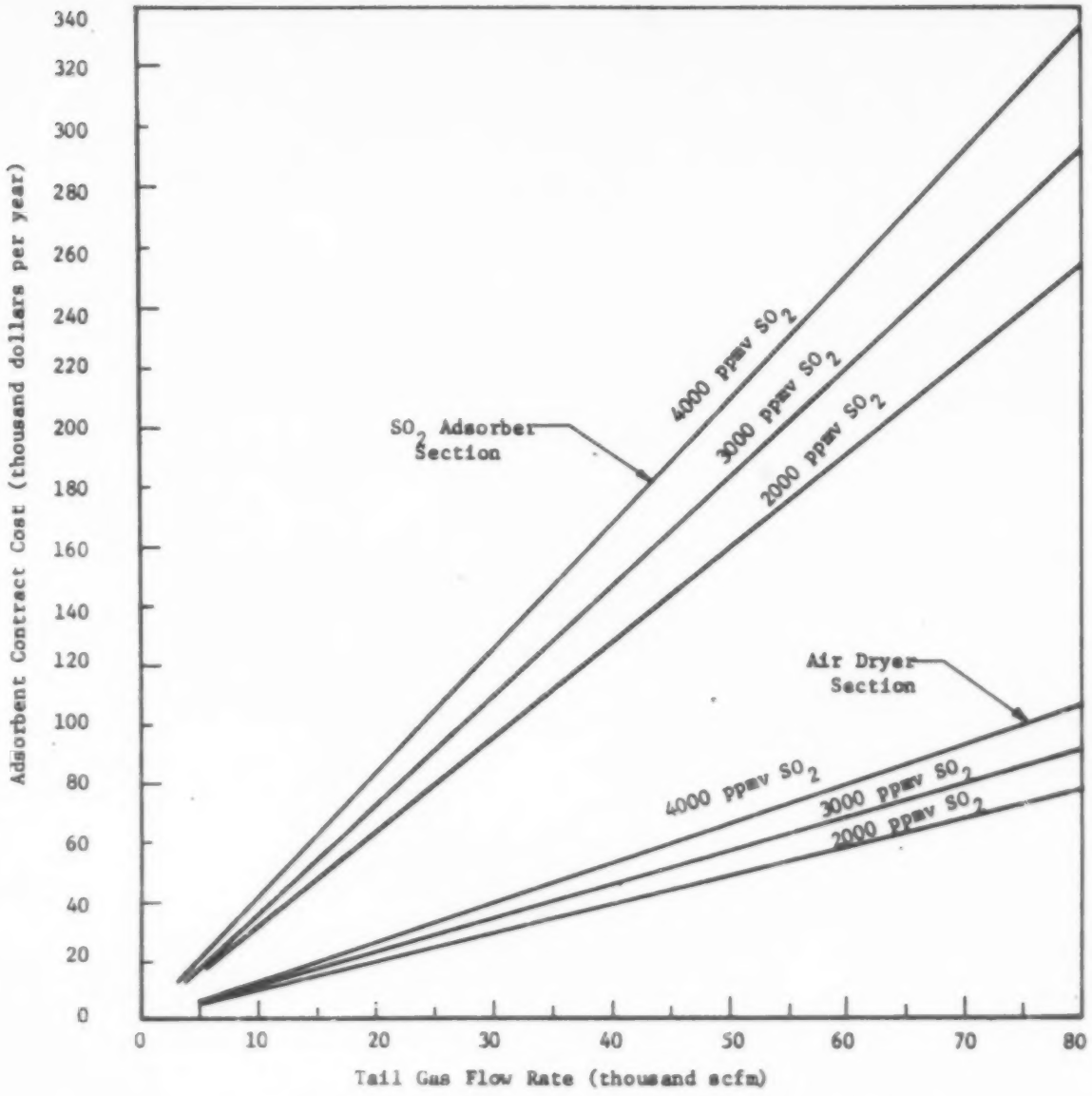

FIGURE B-5. ADSORBENT CONTRACT COST, SIEVE-DRIED REGENERATION, THREE YEARS, 500 PPM SO_2

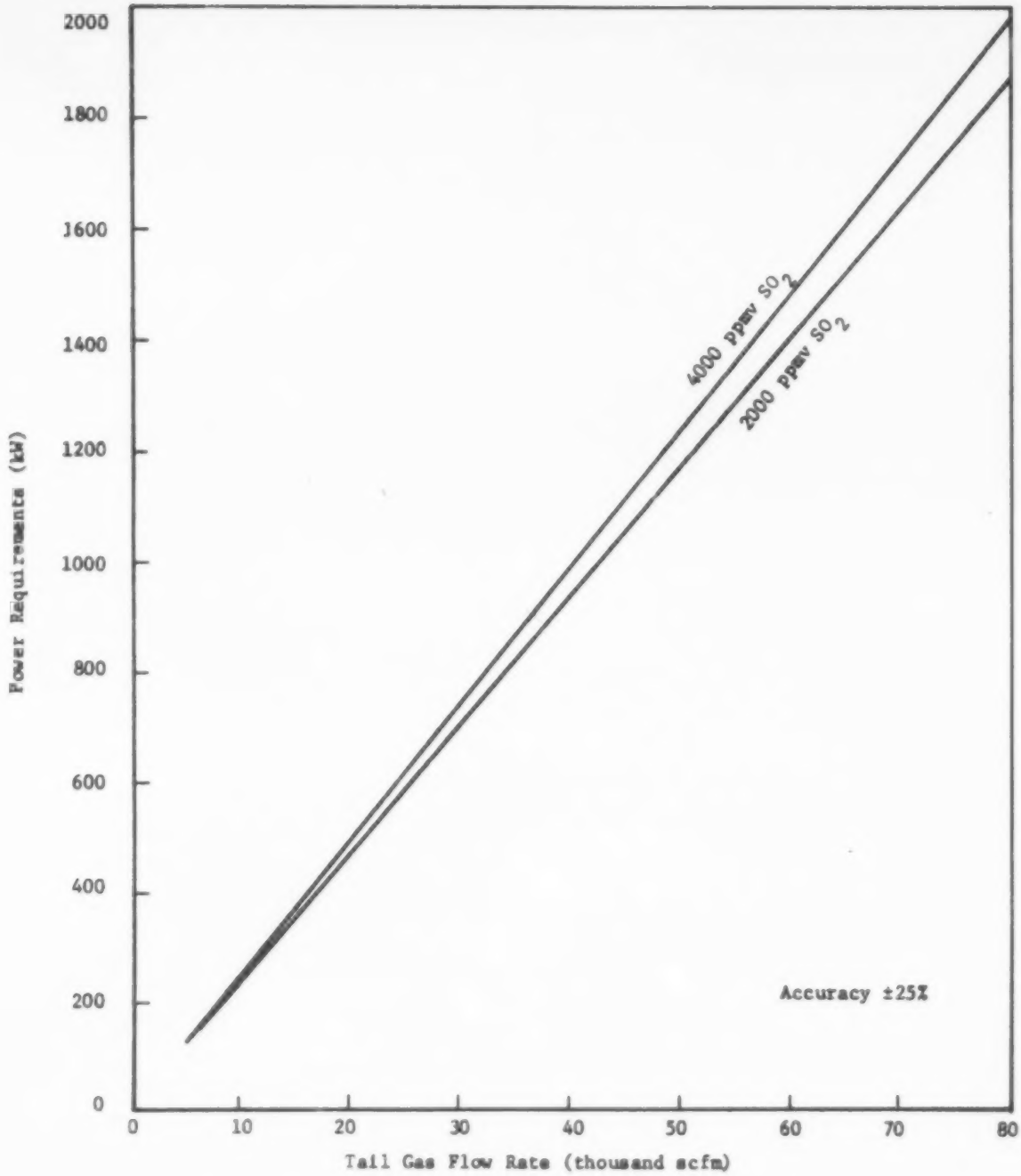

FIGURE B-6. POWER REQUIREMENTS, ACID-DRIED REGENERATION

10/5/73

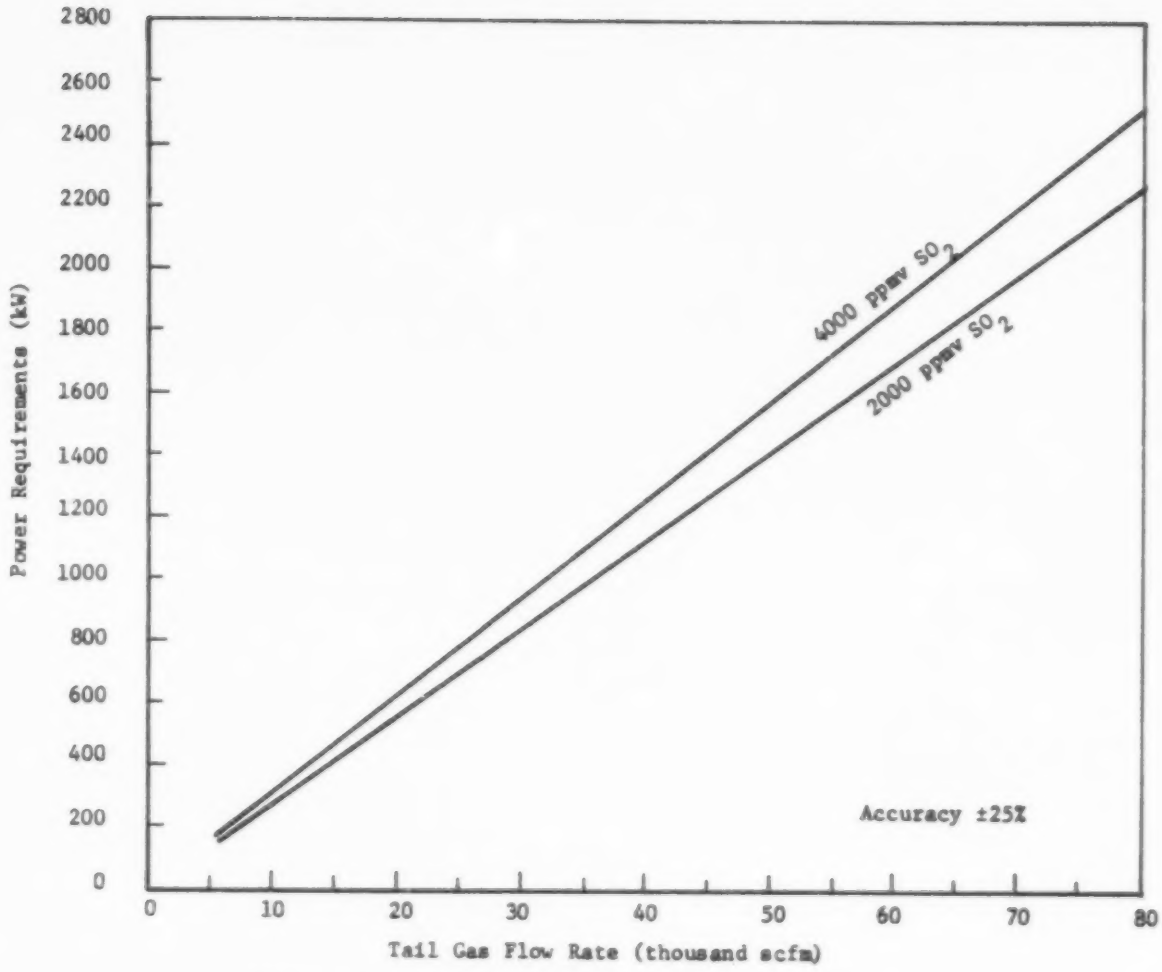

FIGURE B-7. POWER REQUIREMENTS, SIEVE-DRIED REGENERATION

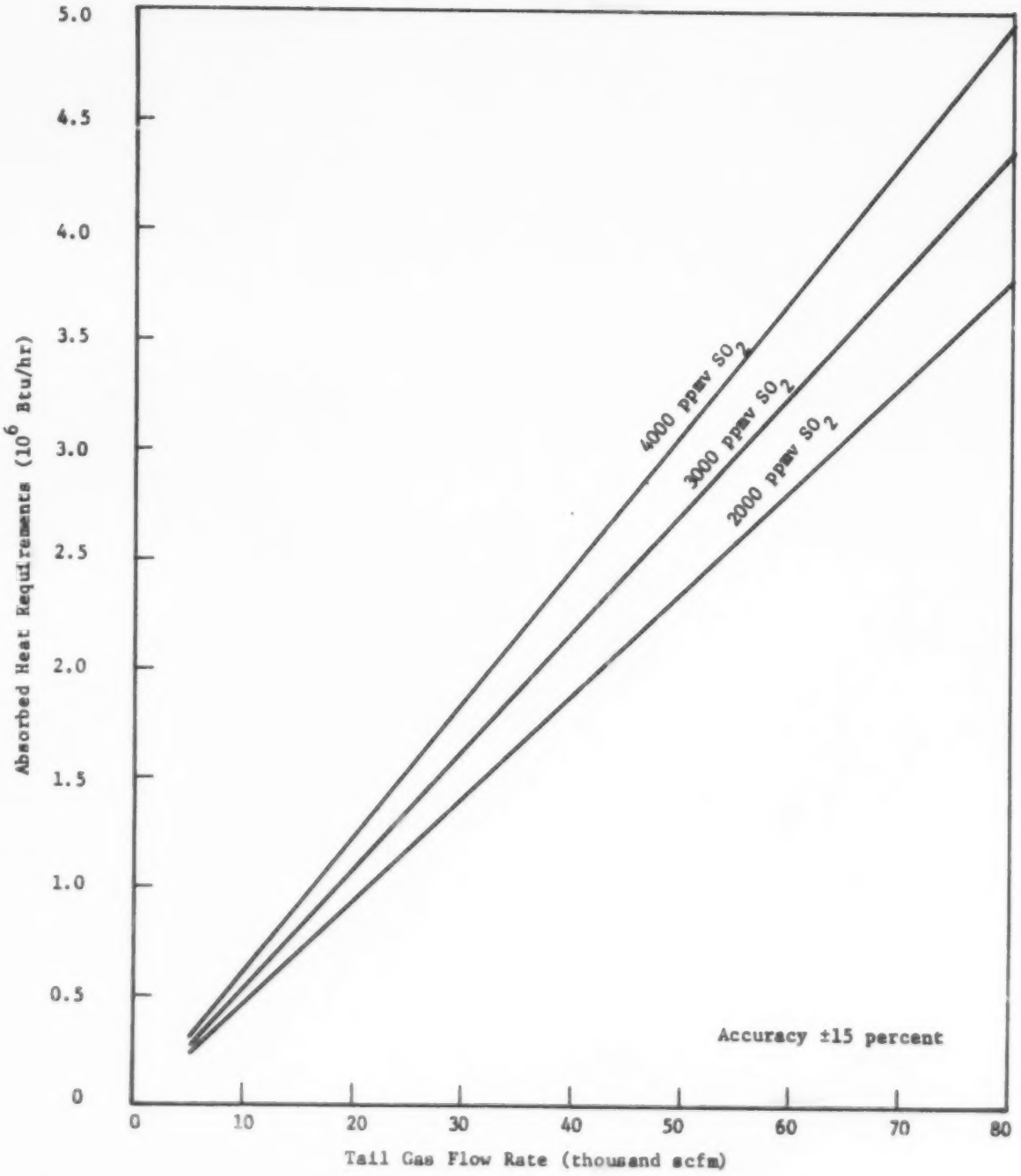

FIGURE B-8. ABSORBED HEAT REQUIREMENTS, ACID-DRIED REGENERATION

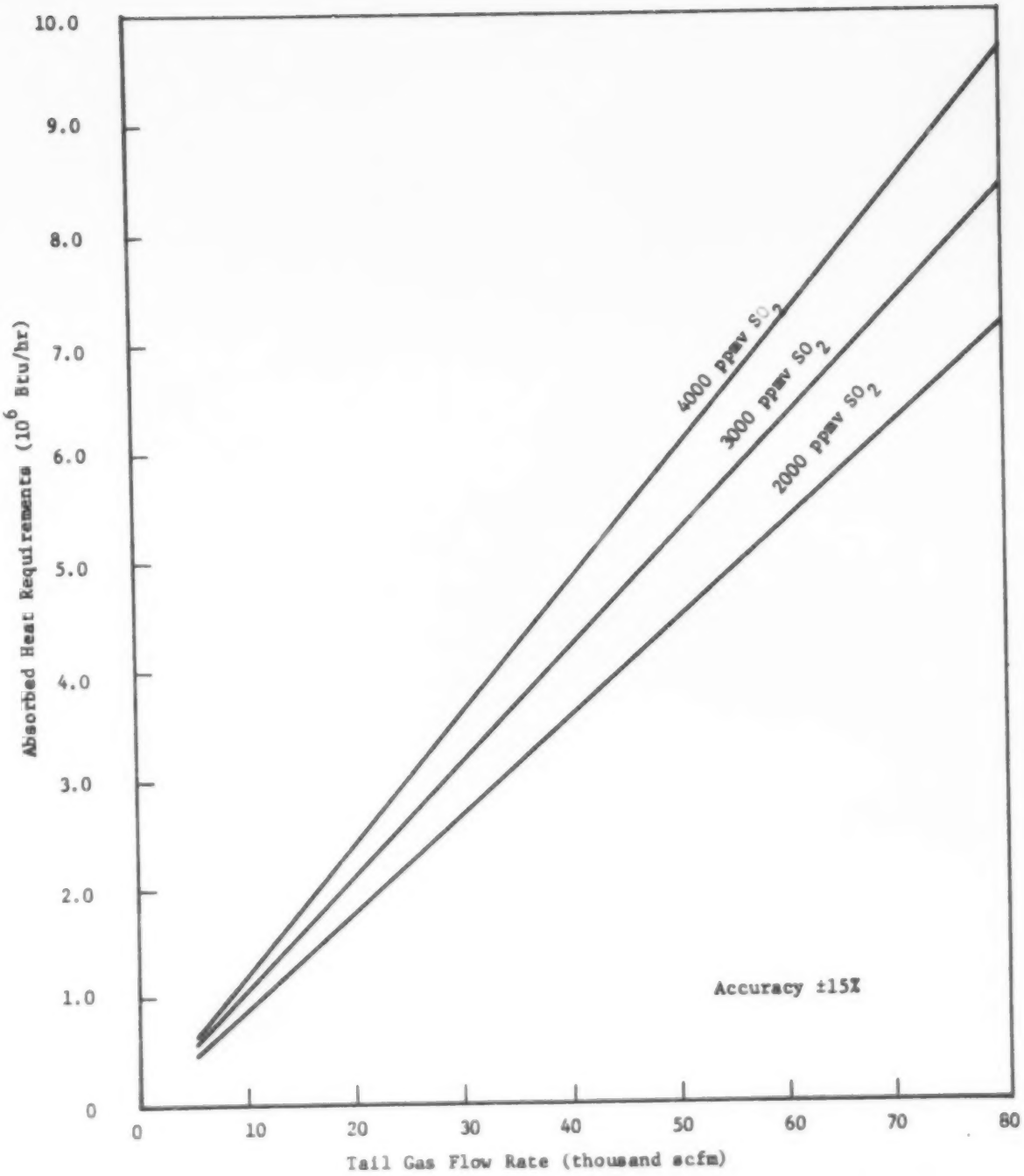

FIGURE B-9. ABSORBED HEAT REQUIREMENTS, SIEVE-DRIED REGENERATION

10/5/73

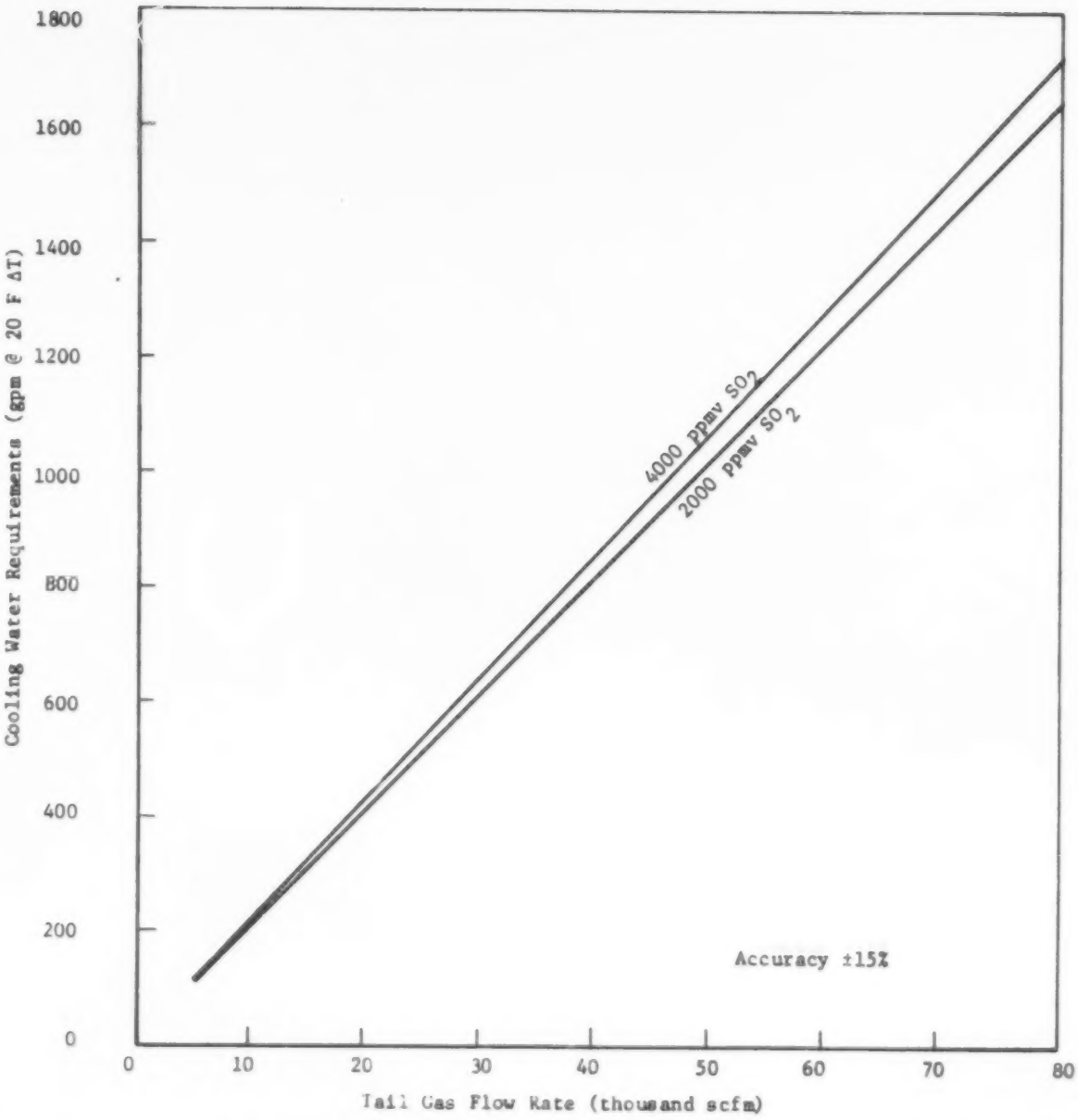

FIGURE B-10. COOLING WATER REQUIREMENTS, ACID-DRIED REGENERATION

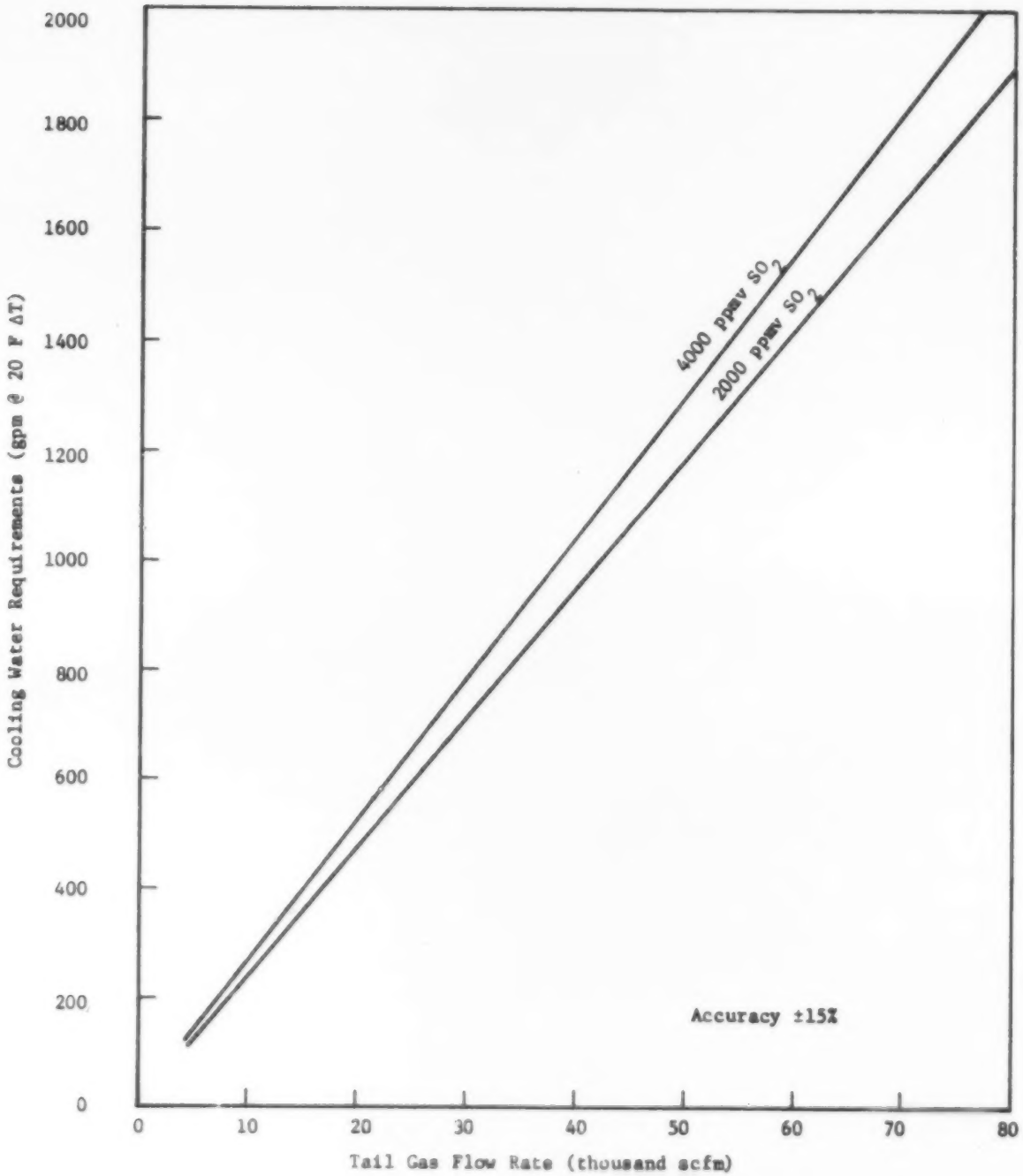

FIGURE B-11. COOLING WATER REQUIREMENTS, SIEVE-DRIED REGENERATION

decreases in the process cost. These additional improvements could not be factored into this study but should be kept in mind when comparisons with other technology options are made.

The Coulton PuraSiv S unit is not a good basis for analyzing the investment required for this process since it is not a typical installation. However, it can be used for an approximate check of the investment level shown in Figure B-1. When the Coulton unit was built in 1972, the investment was about $397,000.[8] This cost was low because

(1) It did not include design, engineering, or contractor's fees since Coulton did their own design and construction work

(2) It did not include the tail gas blower, since Coulton supplied this

(3) Coulton bought used mist eliminators

(4) Union Carbide gave Coulton a low price because this was the first-of-a-kind installation.

The following analysis leads to an approximate 1974 cost for a complete plant. Updating the above cost from 1972 to 1974 using the Marshall and Stevens Index gives a cost of $476,000. A blower designed for the maximum tail gas flow rate of 10,446 scfm, a system pressure drop of 3 psia,* an inlet temperature of 70 F, and an efficiency of 60 percent will require 232 brake horsepower. Based on the cost analysis method of Guthrie,** the 1974 "bare module" cost of this blower would be about $222,000. Adding this to the $476,000 and adding 15 percent for design, engineering, and contractor's fees gives a total investment of about $803,000. This compares reasonably well with the cost of about $900,000 obtained from Figure B-1, the difference being attributable to Items 3 and 4 above.

A rough verification of the electrical power requirements (Figures B-6 and B-7) can also be made from the blower design mentioned above. The 232 brake horsepower for the tail gas blower corresponds to a power requirement of 173 kW. Adding to this the requirements for the regeneration air blowers, using the same 3 psi pressure drop and the air/tail gas flow ratios from Figures 4 and 5 of the text, gives total blower

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* Information obtained by Battelle personnel during visit to Coulton plant on June 12, 1974.

** Guthrie, K. M., "Capital Cost Estimating", Chemical Engineering, p 114, March 24, 1969.

power requirements of 216 kW for an acid-dried air system and 259 kW for a sieve-dried air system. These values are below but fairly close to the requirements of 260 kW (acid dried) and 300 kW (sieve dried) obtained from Figures B-6 and B-7, reflecting the use of some power for items other than blowers.

Wellman-Lord Process

The investment for the Wellman-Lord process was based on the unit installed at Olin Corporation's plant at Curtis Bay, Maryland. The utility and chemical requirements were based on values given by R. T. Schneider of Davy Powergas, Inc.,* since this was the only complete set of requirements available. Davy Powergas is the vendor of the Wellman-Lord process in the United States.

Since data are available from several sources, it may be helpful to compare these data. Table B-1 compares the data from the three sources analyzed as part of this study.

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* Schneider, R. T., "Three Solutions to the SO_2 Emission Problem from Sulfuric Acid Plants", AIChE meeting, Daytona Beach, Florida, May 19, 1973.

TABLE B-1. COMPARISON OF COST DATA FOR WELLMAN-LORD PROCESS

Data Source	Schneider	Olin Curtis Bay Plant	McGlamery & Torstrick
Reference Number	15	13	14
Type of Plant	H_2SO_4	H_2SO_4	Power plant
Plant Size	800 T/D H_2SO_4	1000 T/D H_2SO_4	500 MW
Inlet Gas Rate (10^3 cfm at 60 F)	50.5[a]	67.6	1248
Sulfur Recovery Rate (T/D)	7.98[a]	16.3[b]	112.1
Operating Time (hours/yr)	7992		7000
Directly Derived Data			
Investment (10^3 $)	1,250	2,800	19,184
Date for Investment	May, 1973	Mid-1972	Mid-1974
Electricity (kW)	227	342	
Cooling Water (gal/min)	560		
Steam (lb/hr)	11,000	25,868	305,400[c]
NaOH (T/D)	2.65	5.75	
Other Calculated Quantities			
1974 Investment (10^3 $)[d]	1,452	3,357	19,184
Electricity (kwhr/10^3 scf gas)	0.0751	0.0844	
Cooling Water (gal/10^3 scf gas)	11.1		
Steam (lb/10^3 scf gas)	3.62	6.38	4.08[c]
NaOH (T/T sulfur recovered)	0.332	0.353	

(a) Based on average data for U.S. H_2SO_4 industry, as discussed in text.

(b) Based on quoted average inlet and outlet SO_2 concentrations of 4000 and 293 ppm, respectively.

(c) Includes net steam requirement for SO_2 reduction facility, which could be either positive or negative.

(d) Chemical Engineering Plant Cost Index = 165.4, Marshall & Stevens Index = 398.